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PROCESS DESIGN MANUAL FOR UPGRADING EXISTING WASTEWATER TREATMENT PLANTS

Roy F. Weston, Incorporated
West Chester, Pennsylvania

October 1971

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PROCESS DESIGN MANUAL FOR UPGRADING EXISTING WASTEWATER TREATMENT PLANTS

For
ENVIRONMENTAL PROTECTION AGENCY
Technology Transfer

By

ROY F. WESTON, INC.
Environmental Scientists and Engineers
West Chester, Pennsylvania

Program No. 17090 GNQ
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ABSTRACT

The main purposes of this manual are to examine situations that necessitate upgrading of existing municipal wastewater treatment plants and to discuss and evaluate the corrective actions that are required to upgrade these existing plants. Upgrading to overcome organic and hydraulic overloadings and/or to meet more stringent treatment requirements is considered. The information presented in this manual is specifically adapted to plants having capacities of less than 5 mgd. This particular capacity was selected because most of the existing municipal wastewater treatment plants in the United States have capacities of less than 5 mgd.

The manual emphasizes that operational improvement and modifications to existing unit operations be considered as the logical initial approach to upgrading existing treatment plants, before major expansion of existing facilities is implemented.

Because of the numerous alternatives available for upgrading an existing treatment plant, it is necessary to understand thoroughly the fundamentals of the various unit operations commonly used in municipal wastewater treatment plants. Therefore, this manual examines in depth the capabilities, limitations, and interrelationships of the various unit processes. The manual also examines hypothetical situations requiring upgrading of unit operations and describes "order of magnitude" costs associated with the upgrading of various unit operations.

One chapter of the manual presents case histories of upgrading of existing wastewater treatment plants to illustrate the approaches actually used in these circumstances. The operation and maintenance requirements of the upgraded treatment plants are also briefly examined in the manual.

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FOREWORD

The formation of the Environmental Protection Agency marks a new era of environmental awareness in America. The Agency's goals are national in scope and encompass broad responsibility in the area of air and water pollution, solid wastes, pesticides, and radiation. A vital part of EPA's national water pollution control effort is the constant development and dissemination of new technology for wastewater treatment.

It is now clear that only the most effective design and operation of wastewater treatment facilities, using the latest available techniques, will be adequate to meet the future water quality objectives and to ensure continued protection of the Nation's waters. It is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefit of our pollution control expenditures.

The purpose of this manual is to provide the engineering community and related industry a new source of information to be used in the planning, design, and operation of present and future municipal wastewater treatment facilities. It is recognized that there are a number of design manuals, manuals of standard practice, and design guidelines currently available in the field that adequately describe and interpret current engineering practices as related to traditional plant design. It is the intent of this manual to supplement this existing body of knowledge by describing new treatment methods, and by discussing the application of new techniques for more effectively removing a broad spectrum of contaminants from wastewater.

Much of the information presented is based on the evaluation and operation of pilot, demonstration, and full-scale plants. The design criteria thus generated represent typical values. These values should be used as a guide and should be tempered with sound engineering judgment based on a complete analysis of the specific application.

This manual is one of the four now available through the sponsorship of the Environmental Protection Agency to describe recent technological advances and new information in the following subject areas:

Granular Carbon Adsorption
Phosphorus Removal
Upgrading Existing Plants
Suspended Solids Removal

These manuals are the first edition copies and will be updated as warranted by the advancing state of the art to include new data as it becomes available, and to refine design criteria as additional full-scale operational information is generated.

CHAPTER 1

INTRODUCTION

The ability of wastewater treatment plants to perform at required levels of efficiency becomes more critical as water pollution abatement programs achieve their objectives. Deviations from design performance, which were formerly of lesser consequence, now become paramount because of their impact on the receiving waters. Improved process monitoring and plant operation will obviously reduce the incidence of inadequate performance, but many cases are the result of more basic deficiencies in the treatment system. Such deficiencies can arise from: 1) inadequate initial design; or 2) increased or changed load applied to the system. Another consideration in achieving the required levels of efficiency is the upgraded treatment required to maintain the desired water quality in the receiving waters.

Regardless of the cause, the result is that an inadequately treated effluent is discharged. The historical solution to such a problem has been plant expansion along the same lines as the original facility, or addition of conventional unit processes to add secondary or, in a relatively few cases, tertiary treatment to the system. Depending on its application, a generalized approach such as this does not necessarily make optimum use of the previously existing facilities nor of the expanded facilities. The situation is further complicated where regional treatment systems are proposed for the future and existing facilities are inadequate for the interim period. In such cases, a solution must make optimum use of available technology, with minimum capital expenditure.

Upgrading of wastewater treatment plants may be required to handle higher hydraulic and organic loadings to meet existing effluent quality and/or to meet higher treatment requirements. Any of these situations requires optimization of existing facilities before consideration of additional treatment facilities. It is necessary that a distinction be made between upgrading to accommodate higher hydraulic and organic loads, and upgrading to meet stricter treatment requirements. Existing facilities can be made to handle higher hydraulic and organic loads at slightly reduced treatment efficiency by process modifications, whereas meeting higher treatment requirements usually requires significant expansion and/or modification of existing facilities.

Rapid urbanization, development of industries, and stricter treatment requirements often necessitate unanticipated upgrading of treatment plants or premature implementation of upgrading programs. Many existing treatment plants are not capable of meeting the more stringent performance levels required by today's water quality standards. In addition, there are needs for interim improvements. These considerations, plus economic pressures to optimize pollution abatement expenditures, make it mandatory that a logical and technically sound approach to upgrading existing treatment facilities be established. This is especially true because of the numerous alternatives available for consideration prior to the selection of a method for upgrading an existing facility. It is for this reason that a major plant expansion, i.e., complete duplication of existing unit treatment processes, for the purposes of this manual will be considered the least attractive upgrading procedure available, since this approach does not consider optimization of existing facilities.

Therefore, the purpose of this manual will be to present necessary information for considering various courses of action with regard to an impending or existing plant overload situation, or with regard to increasing the efficiency to meet stricter water quality standards. The diversity of causes that necessitate upgrading of existing plants precludes the use of this manual as a conventional design manual. Therefore, it is aimed at establishing a framework of possible alternative methods of upgrading overloaded treatment plants having capacities up to 5 mgd. This maximum capacity was selected because over 94 percent of the existing treatment plants in the United States in 1968 had capacities of less than 5 mgd (1). Also, past experience has indicated that plants smaller than 5 mgd often have a higher proportion of operational and upgrading problems than do plants of larger size. To facilitate the information presented in the subsequent sections, only plants treating "typical" domestic wastewaters will be considered.

Particular upgrading procedures are stressed as interim methods which may be implemented with a minimum amount of effort and capital expenditure prior to a more elaborate upgrading or even a major plant expansion. Cost information has been compiled and estimates prepared for the upgrading of individual unit processes. When available, cost information has also been included for the reported case histories on plant upgrading. Due to the varying complexity of existing plants, the real benefit of the subsequent cost information will be as a tool for developing comparative capital costs for various upgrading techniques. Particular unit process cost information must be used cautiously, since the complexity of the individual situation will dictate the costs required for upgrading.

The aspects of nutrient removal, although extremely important and oftentimes responsible for upgrading action at many treatment plants, will not be discussed since a separate manual will be published by EPA on this topic.

References

1. *Statistical Summary 1968 Inventory Municipal Waste Facilities in the United States.* Federal Water Quality Administration: Government Printing Office, 1971.

CHAPTER 2

INVESTIGATIVE APPROACH

2.1 Examination of Need for Upgrading

The need for upgrading an existing wastewater treatment plant may arise for one or more of the following reasons:

1. Lack of proper plant operation and control.
2. Inadequate plant design.
3. Changes in wastewater flow or characteristics.
4. Changes in treatment requirements.

2.1.1 Lack of Proper Plant Operation and Control

One of the primary considerations in evaluating an overloaded plant is in the area of improper plant operation and control. An incorrectly operated or maintained plant will never be able to perform according to design. Therefore, no physical upgrading should be considered before the engineer is assured that the plant is being operated to yield its maximum efficiency.

There seem to be two main reasons for the large number of smaller-capacity plants throughout the country which are poorly operated or maintained:

1. The smaller community or sanitary district will not or cannot provide funds for the employment of qualified operators.

2. The lack of appropriation of operating funds limits the extent of any scheduled maintenance program.

In addition, many wastewater treatment plants do not have a laboratory equipped to analyze the wastewater samples from the various units to assess their performance. Improper operation, coupled with inadequate laboratory control, increases the probability of inadequate treatment. For this reason, the smaller community or sanitary district should make sure its plants are staffed with an adequate number of competent operators and laboratory personnel. Further, sufficient funds should be made available to insure a proper maintenance program.

2.1.2 Inadequate Plant Design

In the past, the problems associated with inadequately designed wastewater treatment plants have been a major concern of individual state agencies. For this reason, most states have adopted conservative design guidelines and review procedures which must be followed unless the engineer has operating data which will substantiate a less conservative viewpoint. The implementation of these procedures by regulatory agencies has substantially reduced the problem of inadequate plant design as applied to municipal wastewater.

The major exception to this is in the area of joint municipal and industrial wastewater treatment. Design of wastewater treatment facilities based strictly on state standards may not be applicable when a significant amount of industrial wastewater is discharged to a municipal plant. When this situation occurs, sufficient wastewater sampling and treatability studies should be performed to establish parameters necessary for the design of the treatment plant.

In the past, one of the areas in the design of treatment plants not given much consideration was flexibility. The following design considerations can greatly increase the flexibility allowed to the treatment plant operator:

1. Splitter boxes before and after individual unit processes, for greater ease in operation and maintenance.

2. Piping associated with the aeration basin designed with enough flexibility to facilitate implementation of various activated sludge modifications.

3. Sufficient blower capacity to meet fluctuating organic loads to the aeration basin.

4. Sufficient recycle capacity for trickling filters to meet fluctuating loads.

5. Chlorination capacity with an incremental factor to be utilized for operational control, e.g., odor control.

2.1.3 Changes in Wastewater Flow or Characteristics

Two major problems facing engineers in the design of wastewater treatment plants are: 1) forecasting changes in population and wastewater flow; and 2) the operational problems caused by changes in the characteristics of municipal wastewaters due to the rapid industrialization of an area. Forecasting changes in population and wastewater flows in connection with upgrading of a treatment plant may be quite burdensome, but generally will not be subject to as much uncertainty as in similar forecasting for a relatively undeveloped area. In many cases, the maximum anticipated growth is defined by saturation of the tributary area. Potential extension of this area must also be considered, and is often limited by topographical constraints and political boundaries.

In-line measurement and analysis of existing wastewater flows, analysis of local area growth patterns, examination of local influences such as land use planning studies, zoning regulations, wastewater discharge ordinances, and full use of State, County, and local planning agencies can all be extremely useful in judging the future expected flows for a given upgrading situation.

Rapid industrialization in the plant's service area can cause major operational problems in existing plants, which may require upgrading of various unit processes to handle increased hydraulic and organic loads. The alert community before issuing a building permit should make the industry aware that pretreatment may be required for wastewater containing toxic materials, or for those having an unusually high percentage of organic or inorganic material compared to typical domestic wastewaters. Equalization of industrial wastewater discharges may be helpful in minimizing diurnal flow variations to treatment plants and in distributing shock loads of high-strength wastewaters.

Population equivalent is a reasonable method of equating the organic content or flow contribution of industries to the ordinary per capita contribution present in domestic wastewaters. In many cases, even a relatively small industry may contribute a significantly higher loading than the existing population. Population equivalents for many industrial wastewaters should be based on COD analysis rather than on BOD_5, since extremely strong or toxic wastes may show an artificially low BOD_5 value.

2.1.4 Changes in Treatment Requirements

Increased pressure on the part of the Federal and State governments and a more ecologically minded public are requiring local communities and sanitary districts to enforce existing water quality standards. In addition, many regulatory agencies are stipulating increased organic and solids removal, minimum dissolved oxygen concentration in the plant effluent, a minimum consistent BOD and SS removal during low-flow periods, and removal of substantial portions of the nitrogenous oxygen demand. For example, the Potomac Enforcement Conference has recently required 85 percent removal of nitrogenous oxygen demand from wastewaters treated in the Washington, D.C. metropolitan area (1). Therefore, to meet existing and future requirements, upgrading of existing treatment facilities will often be required.

2.2 Study of Plant Performance History

It is the responsibility of the engineer engaged in upgrading an existing wastewater treatment plant to examine the plant's performance history thoroughly, as the first step in an engineering study. Treatment plant operating records serve two basic functions: 1) providing the operating information necessary for process monitoring and control; and 2) providing the historical record of plant performance. Data collected from a typical plant should include: 1) daily total flow; 2) maximum and minimum flow rates within a day; 3) volume of air added; 4) a series of concentration parameters such as BOD, suspended solids, COD, phosphates, nitrogen, and temperature; 5) consumption of chlorine, coagulants, and neutralizing chemicals; and 6) sludge production.

The engineer should examine all treatment plant records to become familiar with the type of sampling and flow measurement techniques employed by the plant personnel and verify their accuracy. One difficult area in assessing plant performance data is the variation in reliability of influent flow measurement data. Often, plant flow is obtained from some type of flow recording instrument, and reliable flow information is possible only when the treatment plant operator makes it a point to calibrate the flow measuring and recording instruments periodically. A representative portion of the operating data should be evaluated by the engineer. If operational data are not available, then it is the responsibility of the engineer to collect sufficient data to proceed with his upgrading evaluation.

2.3 Identification of Problem Areas

After evaluating the plant's operating records, it is necessary to determine the factors which influence the plant's current performance. There are four problem areas whose effect should be considered in any upgrading situation:

1. Hydraulic and organic overloading.
2. Inadequate organic removal.
3. Inadequate solids removal.
4. Inadequate sludge handling.

The performance of different unit processes within a treatment plant is affected to varying degrees by an increase in hydraulic and organic loading. The relationship between the increases in flow and in organics is also an important consideration. For example, a significant increase in flow without a corresponding increase in organics will generally not be as detrimental as when the increase in flow is also accompanied by a correspondingly large increase in organics.

Nevertheless, the efficiency of most unit operations is affected by hydraulic and organic overloading. The increased flow will increase overflow rates and will decrease available detention time in primary and secondary clarifiers. An increased overflow rate in the primary clarifier will in turn decrease suspended solids and BOD removal at this unit. As a result, solids and BOD loads to secondary treatment processes are increased. Also, hydraulic overloading reduces the compaction of solids that normally takes place in clarifiers and increases the volume of sludge to be handled.

Organic overloading significantly increases the:

1. Organic load per unit of aeration volume.
2. Lbs. BOD applied per pound of MLVSS under aeration.
3. Demand for more oxygen.

These effects significantly decrease the efficiency of activated sludge treatment. A similar effect can be seen in trickling filter plants. In activated sludge plants, an increase in organic loading can reduce the operational stability of the process by causing sludge bulking. Solids removal efficiency in secondary clarifiers is thereby reduced, significantly minimizing the amount of solids that can be carried in the aeration basin. The carryover of biological solids from the final clarifier increases the effluent BOD.

Many treatment plants have anaerobic digestion facilities designed on the basis of a certain volume per capita. Increases in the number of people to be served, together with the increased volume of sludge resulting from poor clarifier performance caused by hydraulic overloading, makes the existing volume in many anaerobic digesters inadequate. This situation creates an operational problem leading to inadequate sludge digestion. The poorly digested sludge further complicates dewatering on vacuum filters or sand beds, because it does not dewater rapidly.

The above discussion clearly indicates that problem areas are interrelated and should be assessed concurrently to determine the logical combination of applicable upgrading procedures. Since optimization of existing facilities is necessary in any upgrading situation, an understanding of various modifications available for the different unit processes is essential. This is the reason for defining the capabilities and limitations of various unit processes in subsequent chapters of this manual.

2.4 Consideration of Applicable Upgrading Techniques

Technology in the field of wastewater treatment in the past decade has provided many innovative upgrading procedures to meet deficiencies in existing processes. Various research projects sponsored by EPA have resulted in a better understanding of various unit processes. In addition, new types of equipment for wastewater treatment have enlarged the range of alternatives available for consideration in upgrading treatment plants.

It has long been recognized that the performance of a wastewater treatment plant is affected by variations in the influent flow. Equalization of extreme flows can dampen the fluctuations in loading to a plant.

Various processes and process equipment are being marketed and successfully used to increase removal in primary and secondary clarifiers. They include the use of chemical coagulation, peripheral-feed clarifiers, and inclined-tube settlers. These procedures, in many cases, have the effect of maintaining good solids removal while maximizing the hydraulic throughput in the existing facilities. Chemical addition in primary and secondary clarifiers can increase solids capture and BOD removal. In addition, several types of screening devices are available as possible substitutes for primary clarification.

Several modifications of the conventional activated sludge process, including step aeration, contact stabilization, and complete mixing, have been adequately studied and have been used to upgrade various treatment plants. A most significant development in the activated sludge treatment process came with the full-scale demonstration of the feasibility and effectiveness of using oxygen aeration as a substitute for air aeration. Plants using oxygen aeration are now being designed, with the single largest plant, at Detroit, Michigan, designed for a flow of 300 mgd.

Another method of upgrading an overloaded secondary plant is to provide additional treatment ahead of the existing biological treatment facilities. The use of plastic media trickling filters should be considered when roughing treatment is indicated. Plastic media filters have been successfully used as roughing filters in industrial wastewater treatment, and it is very likely that they will be used in the future for upgrading of municipal treatment plants.

The true effect of a nitrified effluent on dissolved oxygen in receiving waters has just recently been recognized and substantiated (2). For this reason, some regulatory agencies are requiring nitrification of treatment plant effluents during summer periods, and in some cases are contemplating a year-round nitrification requirement. Nitrification during summer months may be accomplished through modifications to the existing treatment units, such as addition of chemicals to the primary clarifier to decrease the organic loading to existing aeration units. However, dependable year-round nitrification will require a two-stage biological treatment system.

On many occasions, treatment plants which are functioning satisfactorily (design flow not exceeded) are required to improve solids or BOD removal because of more stringent water quality standards. This additional treatment can often be achieved by polishing the treatment plant effluent. Several methods are currently available and have been used successfully, including aerobic and facultative lagoons, microstraining, multi-media filtration, and activated carbon treatment.

Although considerable effort has been made in the study of organic removal processes, the area of sludge handling and dewatering has not received corresponding attention. Inadequate digestion and sludge handling facilities often adversely affect the overall treatment plant operation. The return of supernatant or filtrate from thickening or dewatering units to the head of the plant can impose high oxygen demands on the system and add substantial amounts of fine solids which are difficult to remove in the secondary clarifier. The high concentration of nutrients and organics in such streams and the periodic nature of the return flow often necessitate separate treatment, especially when nutrient removal is a consideration.

Various sludge-handling developments which have been successful are: 1) high-rate anaerobic digestion; 2) aerobic digestion; 3) thickening of sludge prior to digestion to increase the capacity of existing digesters; 4) the use of chemicals to improve thickening and dewatering of sludges; and 5) the use of heat treatment processes for the disposal of sludges. The use of aerobic digestion is likely to alleviate many of the operational problems associated with the anaerobic treatment of sludges.

Having briefly presented a general overview of the technology available for upgrading, the engineer should keep in mind various considerations which will affect the overall economics of upgrading:

1. The physical condition of existing plant equipment and structures as it relates to the use of existing facilities in an upgrading situation.

2. The length of time before a major expansion will be required, based on population and wastewater flow projections.

3. The time required for implementation of various upgrading techniques.

4. Compatability of upgrading procedures with future planned expansions. For example, if the engineer determines that contact stabilization will not work well on a particular wastewater, then step aeration may not be the most economical interim step. Perhaps completely mixed activated sludge or oxygen aeration would be more logical. The reasoning behind this type of decision will be explained further in subsequent chapters.

5. Financial resources available to the community.

6. Costs of the various upgrading techniques that can be used to achieve essentially the same result. The operation and maintenance costs, as well as the capital costs, may be substantially different. Therefore, economic comparison of available alternatives is necessary.

2.5 References

1. *Nitrogen Removal from Wastewaters.* Federal Water Quality Administration, Publication ORD-17010, October, 1970.

2. Courchaine, Robert, *Significance of Nitrification in Stream Analysis - Effects on the Oxygen Balance.* Journal Water Pollution Control Federation, 40, No. 5, pp. 835-847 (1968).

CHAPTER 3

FLOW EQUALIZATION

3.1 General

The cyclic nature of wastewater flows, in terms of volume and strength, is well established. While the concept of flow equalization has been employed in the field of water supply and in the treatment of some industrial wastes, it has not been widely accepted in the municipal pollution control field. Anticipated problems with solids settling, odor, and septicity can be cited as the major factors limiting its use.

Recently, interest in flow equalization for municipal treatment has increased due to the advent of stricter water quality standards, the elimination of plant bypassing, and the increased removal efficiencies that are possible when biological or chemical treatment processes are operated at or near steady-state conditions.

There are two major objectives in the design of flow equalization basins. The first of these is simply to dampen the diurnal flow variations that normally exist in typical municipal wastewater collection systems, and thus achieve a constant or nearly constant flow rate through the downstream treatment processes. In this type of system, little consideration is given to controlling the concentration changes that take place during storage. The major design factors are supplying enough air to keep the basin aerobic and providing adequate mixing to prevent solids deposition. Consideration should be given to locating equalization basins both at the treatment plant site and at strategic upstream locations in the tributary collection system. The upstream locations may offer the added advantage of relieving trunk sewer overload during peak flow periods.

The second objective of flow equalization is to provide the capacity to distribute shock loads of toxic or treatment-inhibiting substances over a reasonable period of time to prevent system failure and to minimize the periodic discharge of harmful contaminants to the receiving stream or surface impoundment. The measurement or estimation of time-dependent concentration profiles and flow-through curves is normally used to analyze the flow characteristics of these systems for determining the effects of tank geometry, effluent weir placement, and mixing regime on changes in contaminant concentrations through the basin.

In all cases, the added costs of flow equalization must be measured against the reduction in downstream process costs and the increased efficiencies that can be achieved by operating these processes under relatively constant loading conditions.

3.2 Determination of Equalization Requirements

To determine the appropriate equalization basin volume, it is necessary to plot an inflow mass diagram of the hourly fluctuations for a typical daily wet-weather wastewater flow.

Figure 3-1 shows the hourly fluctuations for a typical plant. Superimposed on Figure 3-1 is the inflow mass diagram for the hourly flows, the ordinate of which is obtained by accumulating the hourly flows and converting them into equivalent volumes of wastewater.

In Figure 3-1, the slope of line A represents the average rate at which the wastewater is pumped from the equalization basin to the downstream treatment units, which for the particular wet-weather flow in Figure 3-1 is 10,000 gallons/hour. This slope is determined by drawing a straight line through the origin and point C, which is the end of the inflow mass diagram. The maximum required capacity of the equalization basin is determined by drawing lines B and D parallel to line A and tangent to the inflow mass diagram at its maximum and minimum points, E and F. The vertical distance between lines B and D represents the minimum required equalization volume of 30,000 gallons, which is approximately 12.5 percent of the average daily wet-weather flow in this example.

In addition to the volume required to equalize typical wet-weather peaks, the basin must be sized to accommodate any anticipated concentrated plant wastewater streams. Anaerobic digester supernatant and sludge dewatering filtrate are periodically discharged to the front end of the treatment plant and usually have higher organic and nutrient concentrations than typical municipal wastewater. Due to their periodic discharge, these flows create shock load conditions which reduce plant efficiency. COD and ammonia concentrations of 10,000 to 20,000 mg/l and 1,000 mg/l, respectively, are common and consequently create a high oxygen demand. Equalization of these loadings is extremely beneficial to overall plant performance.

The following table is an estimate of the equalization basin's volume requirements for the example shown in Figure 3-1:

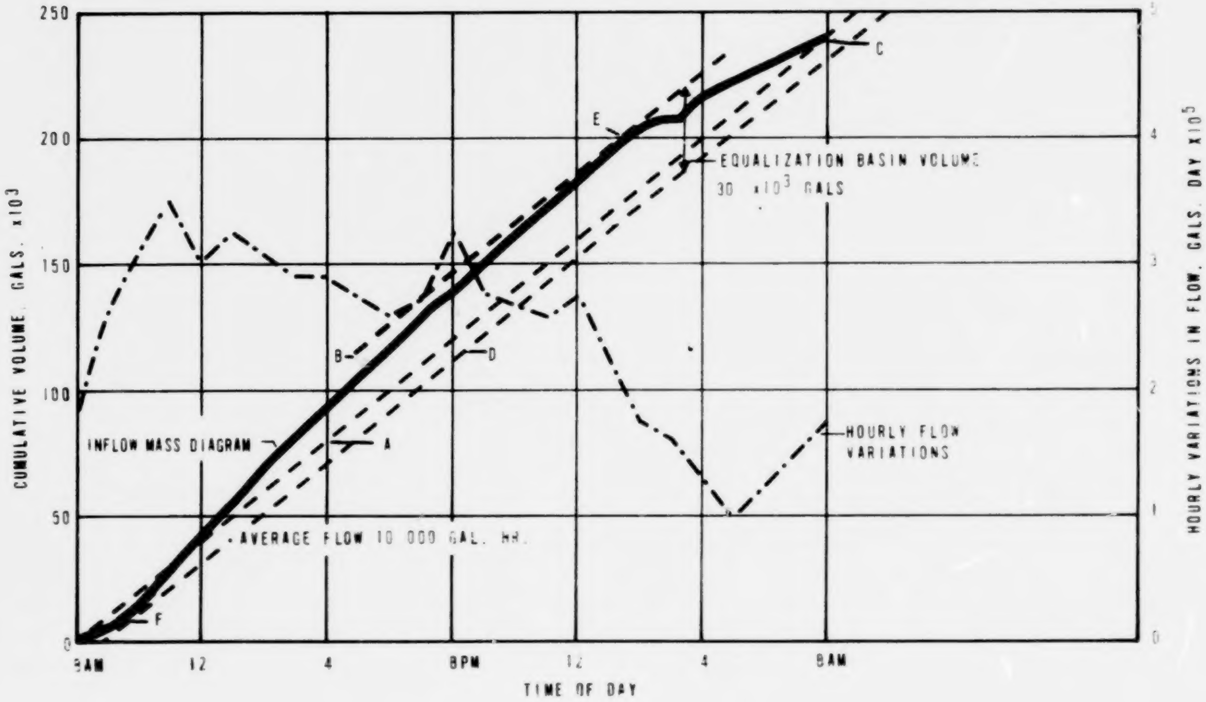
Table 3-1

Equalization Requirements

Source	Range of Equalization Needs percent of flow
Wet Weather Flow Equalization	12.5
Digester Supernatant	0.3 to 1.4
Sludge Dewatering Filtrate	0.5 to 1.5
TOTAL	13.3 to 15.4

The maximum volume requirements to equalize wet-weather flows will depend on the magnitude of the infiltration and extraneous surface water that enters the wastewater collection system. In some cases, it may not be economically feasible to equalize extreme

FIGURE 3-1
EQUALIZATION REQUIREMENTS FOR A TYPICAL WET WEATHER FLOW



peaks of wet-weather flow. Regulatory requirements involving prohibition of treatment plant bypassing, however, will favor the construction of some type of equalization facility. Examination of the plant's past flow records will facilitate the selection of a particular wet-weather flow for design purposes.

For successful operation of equalization basins, mixing and aeration of the wastewater are required. Mixing is necessary to prevent deposition of solids in the basin, and aeration is required to prevent septicity.

A typical equalization flow schematic is shown in Figure 3-2. The flow enters the equalization basin by gravity, and the basin contents are pumped to the primary treatment units using continuous-flow, variable-speed pumps. The maximum pumping capacity of the equalization pump station should be sufficient to handle the maximum flow expected, even though the equalization basin may be designed to equalize a somewhat smaller flow.

3.3 Process Designs and Cost Estimates

Capital cost estimates, prepared for equalization facilities for plants having capacities of 1, 3, and 5 mgd, are shown in Table 3-2. These costs do not include land costs, contingencies, engineering design, and bonding.

Table 3-2

Capital Costs for Equalization Facilities
(ENR Index 1,500)

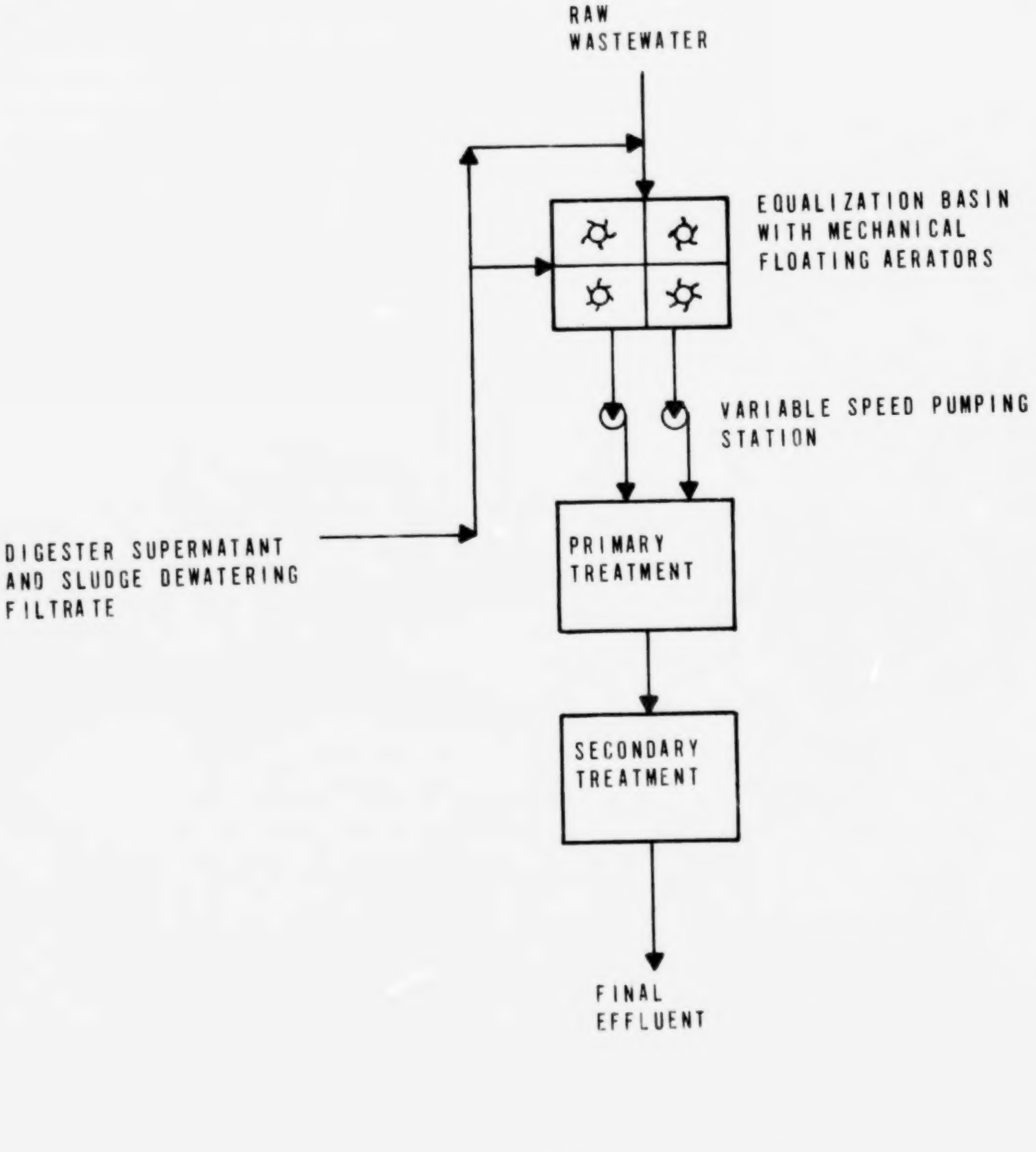
Plant Size mgd	Capital Costs for Equalization Facilities In Thousand Dollars
1	$210
3	450
5	600

The above costs are based on the typical flow diagram shown in Figure 3-2. The volume of the equalization basin was based on 15 percent of the treatment plant daily capacity.

The basin dimensions should be selected to avoid interference between aerators and to minimize fluctuations in basin water level. Aerator manufacturers recommend a minimum basin size of 15 to 50 feet square and a minimum depth of 5 to 8 feet, depending on the particular aerator horsepower used. To avoid large volumes of dead storage in the equalization basin as a result of aerator operating requirements, the use of a compartmented basin is suggested. For example, two compartments of a four-compartment basin could be used to equalize diurnal flow variations, while all four compartments could be used to equalize wet-weather peaks. When floating surface aerators are used, care must be exercised to maintain a minimum water level to protect the aerator. This may be

FIGURE 3-2

SCHEMATIC FLOW DIAGRAM OF EQUALIZATION FACILITIES


RAW WASTEWATER
EQUALIZATION BASIN WITH MECHANICAL FLOATING AERATORS
VARIABLE SPEED PUMPING STATION
DIGESTER SUPERNATANT AND SLUDGE DEWATERING FILTRATE
PRIMARY TREATMENT
SECONDARY TREATMENT
FINAL EFFLUENT

accomplished by compartmentalization as previously mentioned, or by low-level controls on the pump and aerator. Mixing requirements for wastewaters having a suspended solids concentration of approximately 200 mg/l range from 0.02 to 0.04 hp/1,000 gallons of maximum storage volume.

In addition to the mixing requirement, aeration to prevent septicity must also be considered. Oxygen should be supplied to the equalized flow at a sufficient rate (approximately 15 mg O_2/l/hr) to prevent septic odor problems. Mechanical aerators are one method of furnishing both the mixing and aeration requirements. The oxygen transfer capabilities of mechanical aerators operating under standard conditions vary from 3 to 4 lbs O_2/hp-hr.

The costs in Table 3-2 were developed for a reinforced concrete basin equipped with floating aerators. The floating aerator is anchored to the periphery of the basin and is permitted to fluctuate with the water level. The pumps are variable-speed centrifugal pumps. If ground conditions are satisfactory, the use of an earthen lagoon will reduce the cost of the equalization basin significantly.

CHAPTER 4

TECHNIQUES FOR UPGRADING TRICKLING FILTER PLANTS

4.1 General

In 1968 there were more than 3,700 trickling filter plants in the United States serving over 28 million people. In contrast, there were approximately 2,100 activated sludge plants serving 41 million people (1). In the past, the trickling filter plant has been considered the ideal plant for populations of 2,500 to 10,000.

Several reasons have justified this popularity. One is its economy, not only in first cost, but also in operation; another is its relative simplicity of operation, which does not require as highly skilled operators as activated sludge plants require.

Although the effluent from a trickling filter plant is generally of lesser quality than that from an activated sludge plant, trickling filter performance has been considered adequate in many rural areas where stream assimilative capacity is relatively large in relation to population. However, increased urbanization and more stringent water quality standards will require that many existing trickling filter plants be upgraded to improve the quality of treatment provided.

Upgrading of a trickling filter may be required due to hydraulic or organic overloading, higher effluent quality requirements, or both. In general, decreasing hydraulic or organic overloading in existing facilities will not produce a significant increase in BOD removal above the original design value. Therefore, additional treatment facilities will be needed if increased BOD removal is required. It is emphasized that the upgrading of an existing plant should utilize the existing units as much as possible.

4.2 Trickling Filter Processes

Trickling filtration consists of uniform distribution of wastewater over the trickling filter media by a flow distributor. A large portion of the wastewater applied to the filter rapidly passes through it, and the remainder slowly trickles over the surface of the slime. BOD removal occurs by biosorption and coagulation from the rapidly moving portion of the flow and by progressive removal of soluble constituents from the more slowly moving portion of the flow.

The quantity of biological slime produced is controlled by the available food, and the growth will increase as the organic load increases until a maximum effective thickness is reached. This maximum growth is controlled by physical factors including hydraulic dosage rate, type of media, type of organic matter, amount of essential nutrients present, and the nature of the particular biological growth.

In the past, trickling filters have been classified as either low (standard), intermediate, high, or super-rate filters based on hydraulic and organic loading rates.

4.2.1 Low-Rate Trickling Filters

Low-rate trickling filters are designed to handle organic loadings of 10 to 20 lbs. of BOD/1,000 cu.ft./day, and hydraulic loadings of 2 to 4 million gallons/acre/day (mgad). In general, low-rate filters do not use recirculation to maintain a constant hydraulic loading, but use either suction-level controlled pumps or a dosing siphon. Dosing tanks are small, usually with only 2-minute detention time based on twice the average design flow so that intermittent dosing is minimized. Even so, at small plants the low night-time flows may result in intermittent dosing. If the interval between dosings is long (e.g., greater than one or two hours), the efficiency of the process will be affected since the character of the biological slime will be altered due to lack of moisture.

Under normal operations, the low-rate filter and secondary clarifier may average 85 percent BOD removal. By the addition of recirculation during periods of low flows (so that the filter is always wet), it is possible to increase filter efficiency to 90 percent and even higher in some instances (2).

In most low-rate filters, only the top 2 to 4 feet of the filter media have appreciable biological slime. As a result, the lower portions of the filter may be populated by autotrophic nitrifying bacteria which oxidize ammonia nitrogen to nitrite and nitrate forms. If the nitrifying population is sufficiently well established and if climatic conditions are favorable, a well operated low-rate filter, in addition to providing good BOD removal, can produce a highly nitrified effluent. The positive effect that a nitrified effluent has in reducing the total oxygen demand in receiving waters is being increasingly utilized in the formulation of water quality standards.

4.2.2 Intermediate-Rate Trickling Filters

Intermediate-rate trickling filters are generally designed to treat hydraulic loadings of 4 to 10 mgad and corresponding organic loadings ranging from 15 to 30 lbs. BOD/1,000 cu.ft./day, including recirculation. In the past, there have been some cases where the organic loading in the intermediate range stimulated considerable biological filter growth and the rate of hydraulic loading was not sufficient to eliminate clogging of the trickling filter media (2). This clogging situation can be remedied somewhat by utilizing relatively large stone, 3 to 4 inches in diameter. However, it should also be noted that many plants operate in this intermediate range with no reported operational problems (2) (3). In practice, some engineers will design a high-rate filter to operate in the intermediate range during the early period of its operating life, when average flows are substantially below the average design flows.

4.2.3 High-Rate Trickling Filters

High-rate trickling filters have hydraulic loadings of 10 to 30 mgad and organic loadings up to 90 lbs. BOD/1,000 cu.ft./day, including recirculation. In all high-rate filters, some form of recirculation is used in order to maintain a relatively constant hydraulic loading. The correspondingly higher loadings result in an overall BOD removal efficiency that is somewhat lower than that obtainable from a low-rate trickling filter. The higher organic loadings in high-rate filters preclude the development of nitrifying bacteria in the lower section of the filter. Hence, these plants will seldom exhibit any incipient nitrification.

4.2.4 Super-Rate Trickling Filters

Super-rate trickling filters have evolved as a result of the development of various types of synthetic trickling filter media. Past experience has indicated that hydraulic loadings of 150 mgad and higher, including recirculation, may be accommodated in super-rate trickling filters. A discussion of synthetic media characteristics is presented in Section 4.3.2.

4.3 Trickling Filter Performance Factors

There are numerous factors that affect the performance of trickling filters. Some of these are:

1. Wastewater Characteristics.
2. Trickling Filter Media.
3. Trickling Filter Depth.
4. Recirculation.
5. Hydraulic and Organic Loading.
6. Ventilation.
7. Temperature of Applied Wastewater.

4.3.1 Wastewater Characteristics

Domestic wastewaters vary in composition and strength, depending on the relative amounts of industrial wastewater and infiltration present. The rate of BOD removal from a domestic wastewater in a trickling filter generally exceeds the BOD removal rate from an industrial wastewater which has a high percentage of dissolved BOD. This is due to the high percentage of colloids in domestic wastewater, and to the apparent increased ability of the filter to remove this colloidal material. A reasonable explanation for this is that some of these materials are removed by biological flocculation and not by oxidation and synthesis of new cells.

The strength of wastewaters can vary substantially over a daily period. One method of dampening these fluctuations is to recirculate filter effluent through the primary clarifier.

4.3.2 Trickling Filter Media

The introduction of synthetic media for trickling filters has extended the range of hydraulic and organic loading well beyond the range of stone media. Table 4-1 presents a comparison of physical properties of various types of trickling filter media. Two properties which are of interest are specific surface area and percent void space. Greater surface areas permit a larger mass of biological slimes per unit volume, while increased void space allows for higher hydraulic loadings and enhanced oxygen transfer. The ability of synthetic media to handle higher hydraulic and organic loadings is directly attributed to the higher specific surface area and void space of these media compared to stone media and blast furnace slag, as shown in Table 4-1.

Table 4-1

Comparative Physical Properties of Trickling Filter Media

Packing	Nominal Size inches	Units per cu.ft.	Unit Weight lbs./cu.ft.	Specific Surface Area sq.ft./cu.ft.	Void Space percent
Plastic Media	20 x 48	2-3	2-6	25-30	94-97
Del-pak Redwood Media	47½ x 47½ x 35¾	–	10.3	14	----
Granite	1-3	–	90	19	46
Granite	4	–	--	13	60
Blast Furnace Slag	2-3	51	68	20	49

4.3.3 Trickling Filter Depth

Most low-rate trickling filters are designed with depths ranging from 5 to 7 feet, while high-rate filters are designed with depths of 3 to 6 feet. The relatively deep low-rate filters improve the nitrification potential.

The treatment efficiency of a synthetic media trickling filter is much more responsive to variations in depth than a stone media trickling filter. For this reason, the depth selection for a super-rate trickling filter is a major design parameter, as will be illustrated later in this chapter.

4.3.4 Recirculation

The practice of effluent recirculation can be used to improve the efficiency and operation of stone media trickling filters. For example, it can minimize the operational problems associated with intermittent dosing of low-rate trickling filters. Recirculation ratios of 0.5 to 4.0 have been used in high-rate filters; Galler and Gotaas (4) have demonstrated that a recirculation ratio of greater than 4 does not materially increase the efficiency of filters and is also uneconomical.

There are many possible flow configurations which may be used with a single or two-stage high-rate trickling filter plant. Some of the more common flow diagrams which have been presented in the Water Pollution Control Federation's Sewage Treatment Plant Design Manual (MOP No. 8) have been included in Figure 4-1.

Decisions regarding the use of any one of the flow configurations shown in Figure 4-1 are based on an examination of the relative economics and, in some cases, the preferences of the design engineer.

Recirculation as applied to the plastic media involves a slightly different concept than has been previously applied to stone filters. Various types of super-rate filter media have different minimum wetting rates, i.e., a rate of flow per unit area which will induce a biological slime throughout the depth of the media. This minimum wetting rate typically ranges from 0.5 to 1.0 gpm/sq.ft., depending on the geometric configuration of the media. Therefore, recirculation in plastic media filters is practiced to maintain the desired wetting rate for a particular medium. Generally, increasing the hydraulic loading substantially above the minimum wetting rate decreases the BOD removal through the filter (5).

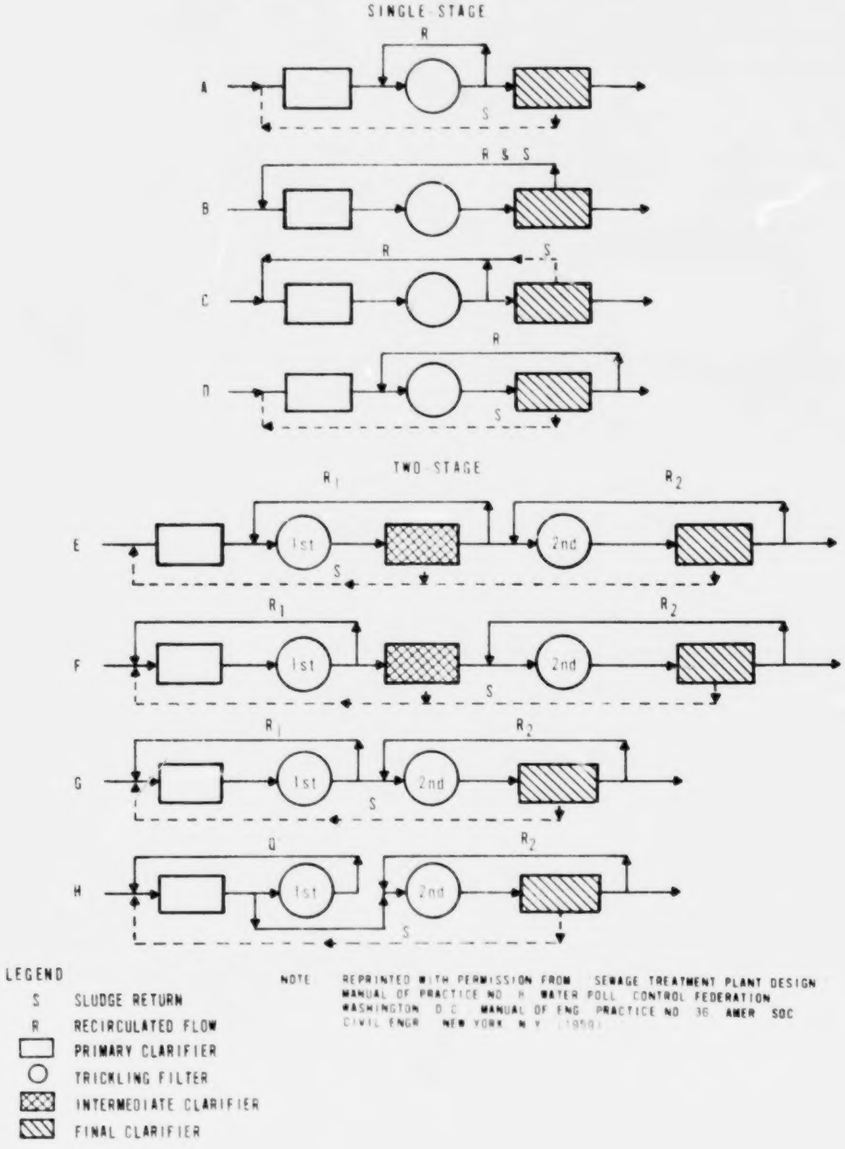
4.3.5 Hydraulic and Organic Loading

Two major parameters which affect the performance of a trickling filter are its hydraulic and organic loading rates. An attempt was made to correlate the efficiency of the secondary portion of various trickling filter plants to their corresponding hydraulic and organic loading rates. The results are shown in Figures 4-2 and 4-3 for stone media trickling filter plants having various recycle rates. It is clear from Figures 4-2 and 4-3 that hydraulic loading will more closely predict the performance of a stone media trickling filter than organic loading. A similar effect is obtained for plastic media trickling filters as shown in Figures 4-4 and 4-5.

4.3.6 Ventilation

Proper ventilation of trickling filters is essential to the maintenance of aerobic conditions throughout the filter media. The Ten-States Standards recommend that all drains, channels, and pipes be sized such that not more than 50 percent of their cross-sectional area will be submerged at the design hydraulic loading (15). If the trickling filter is constructed on or near grade, provision for ventilation will be less critical than if the topography necessitates construction well below grade. In these latter instances, forced ventilation or ventilation shafts may be a consideration. However, many design engineers are of the opinion that forced ventilation is generally not justified (3).

FIGURE 4-1
COMMON FLOW DIAGRAMS FOR SINGLE AND TWO-STAGE HIGH-RATE TRICKLING FILTERS


SINGLE-STAGE
A
R
S
B
R & S
C
R
S
D
R
S
TWO-STAGE
E
R_1
R_2
1st
2nd
S
F
R_1
R_2
1st
2nd
S
G
R_1
R_2
1st
2nd
S
H
Q
R_2
1st
2nd
S
LEGEND
S SLUDGE RETURN
R RECIRCULATED FLOW
PRIMARY CLARIFIER
TRICKLING FILTER
INTERMEDIATE CLARIFIER
FINAL CLARIFIER

NOTE: REPRINTED WITH PERMISSION FROM "SEWAGE TREATMENT PLANT DESIGN," MANUAL OF PRACTICE NO. 8, WATER POLL. CONTROL FEDERATION, WASHINGTON, D.C., MANUAL OF ENG. PRACTICE NO. 36, AMER. SOC. CIVIL ENGR., NEW YORK, N.Y. (1959).

FIGURE 4-2

EFFECT OF HYDRAULIC LOADING ON STONE MEDIA TRICKLING FILTER PERFORMANCE

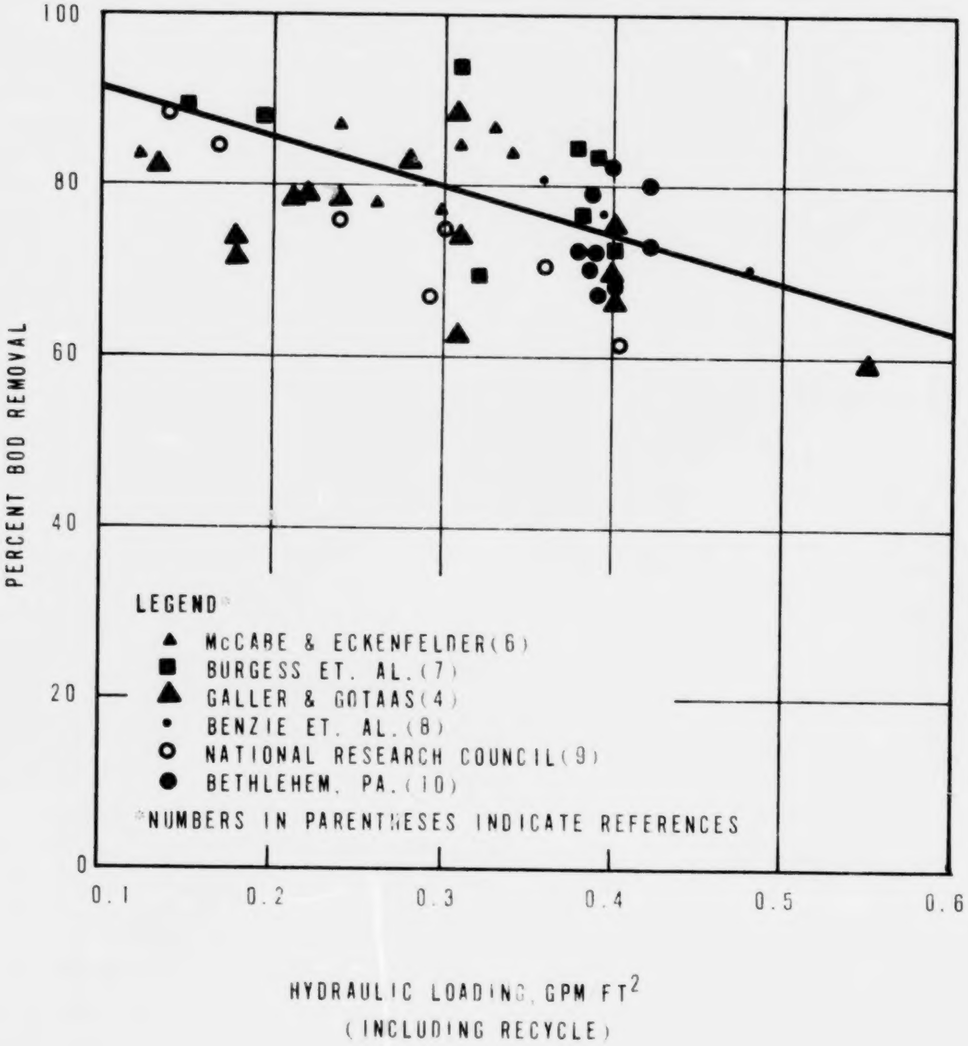


FIGURE 4-3

EFFECT OF ORGANIC LOADING ON STONE MEDIA TRICKLING FILTER PERFORMANCE

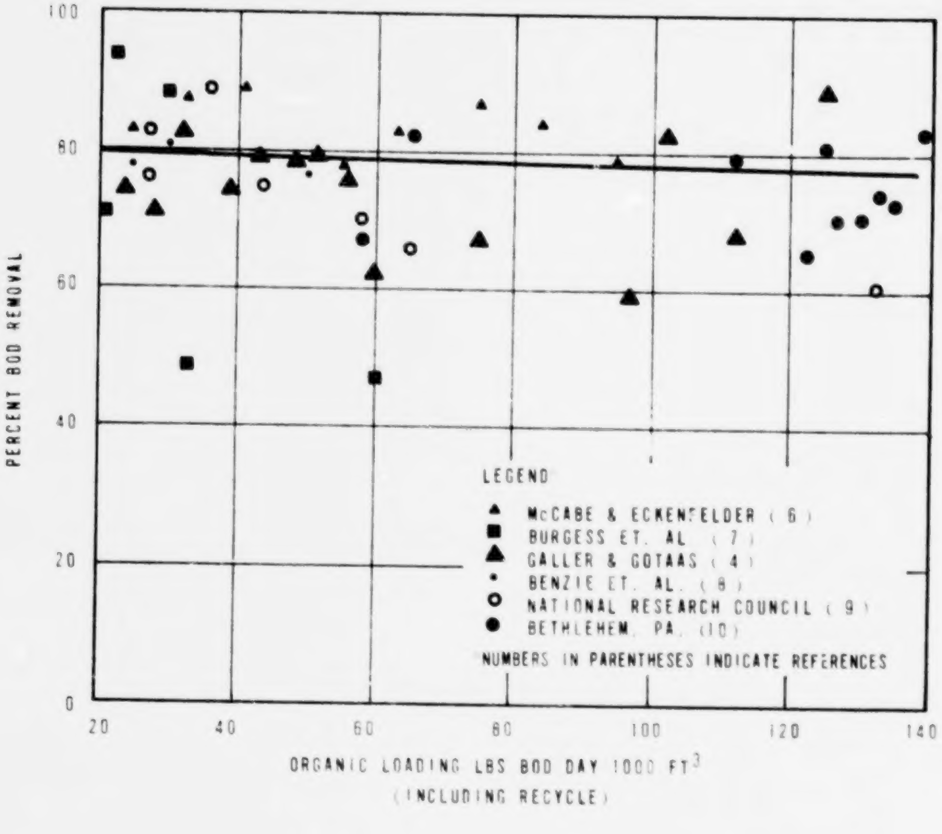


FIGURE 4-4

EFFECT OF HYDRAULIC LOADING ON PERFORMANCE OF PLASTIC MEDIA TRICKLING FILTERS

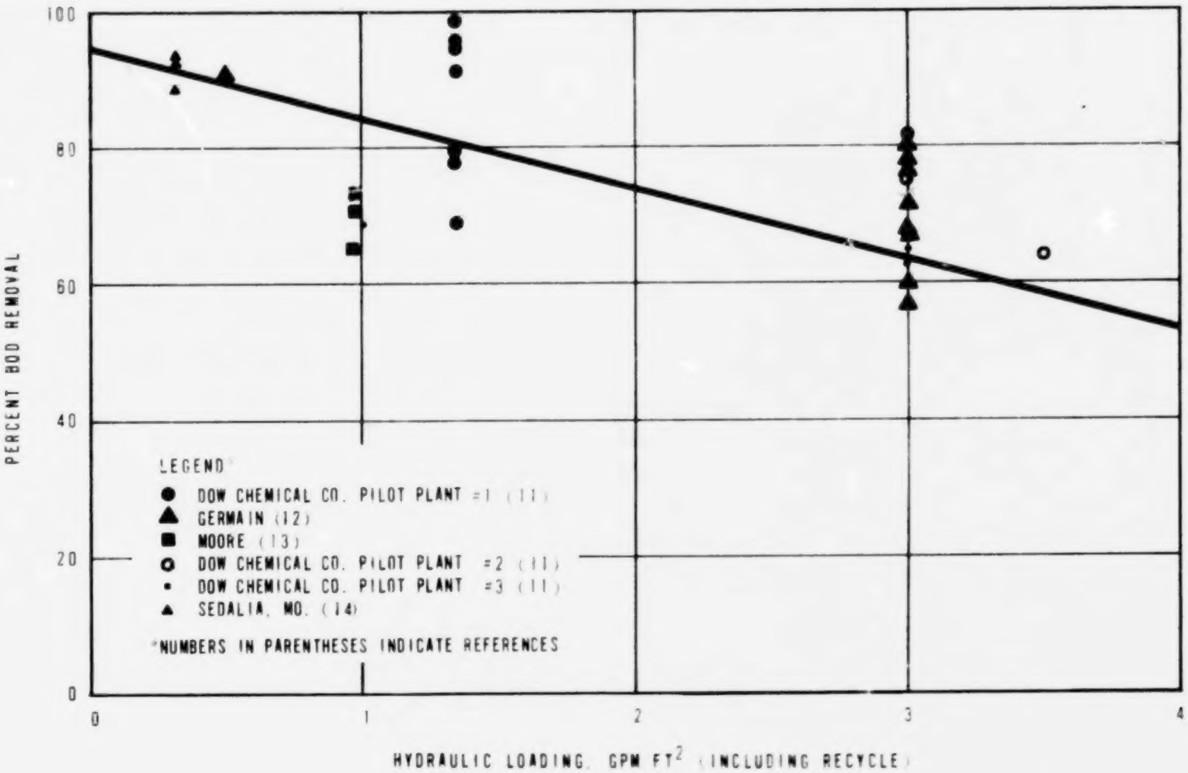
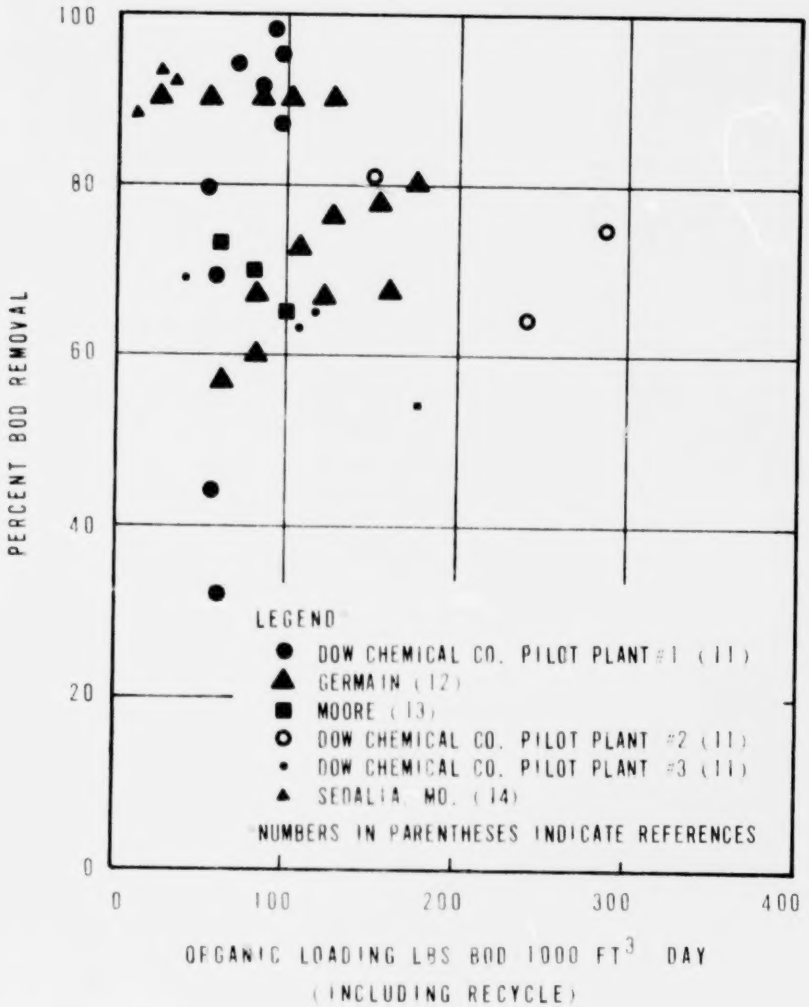


FIGURE 4-5

EFFECT OF ORGANIC LOADING ON PERFORMANCE OF PLASTIC MEDIA TRICKLING FILTERS



4.3.7 Temperature of Applied Wastewater

The efficiency of trickling filters is affected by temperature changes. The effect of temperature on filter performance is expressed by the following relationship (16) (17):

$$E_T = E_{20}\theta^{T-20}$$

where:

θ = Constant varying from 1.035 to 1.041
E_T = Filter efficiency at temperature, T
E_{20} = Filter efficiency at 20°C
T = Wastewater temperature, °C

Filter performance was observed to vary 21 percent between summer and winter months in several high-rate filters in Michigan (8). The effect of temperature was especially pronounced in high-rate filters due to the cooling effect of recirculation. It has been reported that covering of filters in cold climates does not substantially increase the performance because the filter covering does not increase the temperature of the applied wastewater (18).

4.4 Trickling Filter Design Formulas

Several attempts have been made to delineate the fundamentals of the trickling filter process based on actual operating data from trickling filter plants correlating several variables that affect trickling filter operation. Analyses of operating data were made to establish equations or curves that best fitted the available data. The results of these data analyses led to the development of the following various trickling filter formulations:

1. National Research Council Formula.
2. Ten-States Standards.
3. Velz.
4. Rankin.
5. Galler and Gotaas.
6. Schulze.
7. Eckenfelder.

Although the trickling filter formulas represent attempts to include many of the variables that can affect trickling filter operations, the use of any one of these formulas does not universally reflect the actual performance of filters.

4.4.1 National Research Council Formula (NRC) (9)

The NRC formulation was the result of an extensive analysis of operational records from stone-media trickling filter plants serving military installations. Based on data analysis, the NRC recommended the following formulas for predicting the performance of stone-media trickling filters:

First or Single Stage:

$$E_1 = \frac{100}{1 + 0.0085\left(\frac{W}{VF}\right)^{1/2}}$$

Second Stage:

$$E_2 = \frac{100}{1 + \frac{0.0085}{1 - E_1}\left(\frac{W^1}{VF}\right)^{1/2}}$$

where:

E_1 = Percent BOD removal efficiency through the first-stage filter and clarifier
W = BOD loading (lbs./day) to the first or single-stage filter, not including recycle
V = Volume of the particular filter stage in acre-ft.
F = Recirculation factor for a particular stage, $(1 + R)/(1 + 0.1R)^2$
R = Recirculation ratio = recirculated flow/plant influent flow
E_2 = Percent BOD removal efficiency through the second-stage filter and clarifier
W^1 = BOD loading (lbs./day) to the second-stage filter, not including recycle

Some of the limitations of the NRC formulas are:

1. Military wastewater is characteristically more concentrated than average domestic wastewaters.

2. The effect of temperature on trickling filter performance is not considered.

3. NRC formulas indicate that organic loading has a greater influence on filter efficiency than hydraulic loading. This is probably because of the concentrated nature of the wastewaters.

4. Applicability is limited to concentrated domestic wastewaters because no factor is included to account for differing treatability rates.

5. The formula for second-stage filters is based on the existence of intermediate settling tanks following the first-stage filters.

A comparative plot of trickling filter operational data with the predicted value using the first or single-stage NRC formula is shown in Figure 4-6. It is clear from Figure 4-6 that the use of the NRC formula may result in substantial deviation from the actual performance of a trickling filter.

4.4.2 Ten-States Standard Design Guidelines

The data analysis of plants located in the colder northern regions of the United States by the Great Lakes-Upper Mississippi River Board of Sanitary Engineers led to the development of design guidelines for trickling filters. In the 1968 edition of the Ten-States Standards, the Board has presented a loading curve for single-stage stone media filters which is reproduced in Figure 4-7 (15). In developing Figure 4-7, loading due to recirculation has not been considered.

The limitations of this design curve are:

1. The formulation is based on data obtained from colder regions.
2. Hydraulic loadings are considered to have no influence on the efficiency of the filter.
3. Applicability is limited to domestic wastewaters within a specific concentration range.

4.4.3 Velz Formula (21)

In 1948, Velz proposed the first major formulation delineating a fundamental law as contrasted to previous attempts based on data analysis. The Velz formula relates the BOD remaining at depth D as follows:

$$\frac{L_D}{L} = 10^{-KD}$$

where:

L = Total removable BOD, mg/l
L_D = Removable BOD at depth D, mg/l
D = Filter depth, ft.
K = Constant

Removable BOD in the Velz formula is defined as the maximum fraction of applied BOD removed at a specific hydraulic loading range.

FIGURE 4-6

COMPARISON OF TRICKLING FILTER OPERATING DATA WITH NRC FORMULA

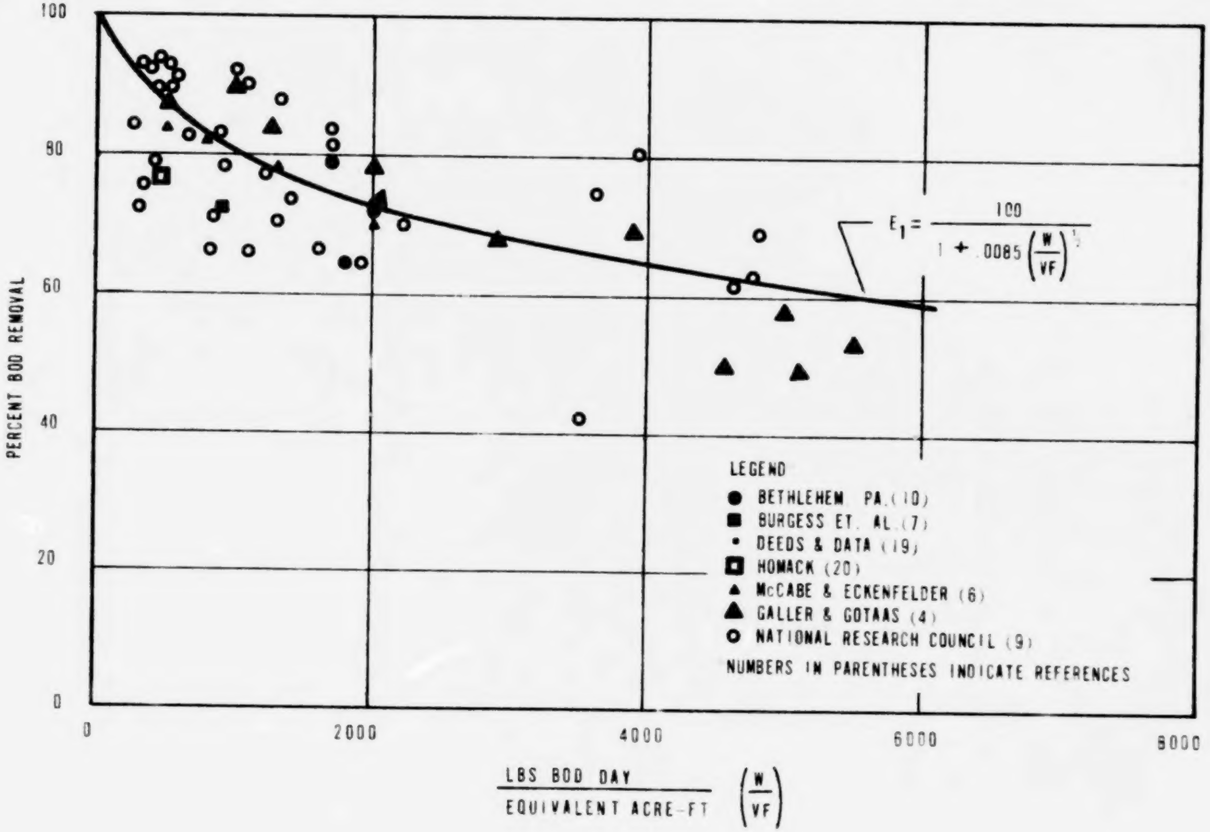
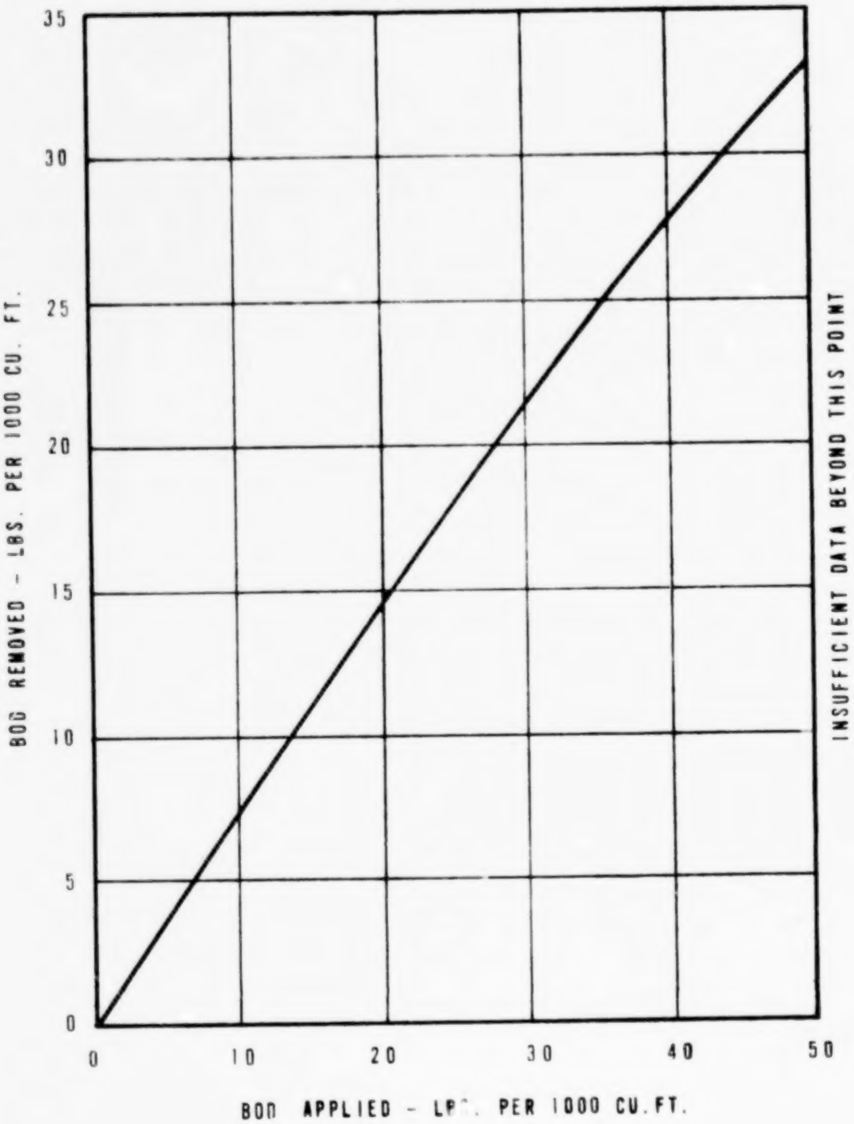


FIGURE 4-7

TEN STATE STANDARD DESIGN GUIDELINE (15) *



*INCLUDE BOD REMOVAL IN THE TRICKLING FILTER AND SECONDARY CLARIFIER.

4.4.4 Rankin Formula

In 1955, Rankin developed empirical formulas based on the Ten-States Standards, including the following equation for a single-stage plant (6):

$$L_e = \frac{L_a}{2R+3}$$

where:

L_e = BOD of settled filter effluent, mg/l
L_a = BOD of primary effluent, mg/l
R = Recirculation ratio

For two-stage, high-rate filters, the Ten-States Standards recommends that "the BOD load applied to the second-stage filter, recirculation included, shall not exceed two times the BOD expected in the settled effluent. When the effluent of the first-stage filter is applied directly to the second stage without intermediate settling, the assumed BOD removal by the first stage shall not exceed 50 percent.(6)" Based on the previous statements, Rankin developed the following equations:

$$L_{e1} = 0.5\ L_a$$

and

$$L_{e2} = \frac{L_{e1}}{R_2 + 2}$$

where:

L_a = BOD of primary effluent, mg/l
L_{e1} = BOD of the unsettled effluent of first-stage filter, mg/l
L_{e2} = BOD of the settled effluent of second-stage filter, mg/l
R_2 = Recirculation ratio of second-stage filter

4.4.5 Galler and Gotaas Formula

In 1964, the last major effort to forecast the performance of stone filters was attempted by Galler and Gotaas (4) using multiple regression analysis of data from existing plants.

Based on regression analysis, the following equation was developed:

$$L_e = \frac{K\,(i\,L_i + r\,L_e)^{1.19}}{(i + r)^{0.78}(1 + D)^{0.67}a^{0.25}}$$

where:

$$K = \frac{0.464\left(\frac{43{,}560}{\pi}\right)}{i^{0.28}T^{0.15}}$$

L_e = Unsettled filter effluent BOD, mg/l
L_i = Filter influent BOD, mg/l
D = Filter depth, feet
i = Influent flow, mgd
r = Recirculation flow, mgd
a = Filter radius, feet
T = Wastewater temperature, °C

The Galler and Gotaas formula recognized the effects of recirculation, hydraulic loading, filter depth, and wastewater temperature as being important in understanding the performance of a trickling filter. They further indicated that recirculation improves the performance of a filter, but established a 4:1 ratio as a practical upper limit for recirculation.

4.4.6 Schulze Formula

In 1960, Schulze (22) postulated that the time of liquid contact with the biological mass is directly proportional to the filter depth and inversely proportional to the hydraulic loading rate; this is expressed as follows:

$$t = \frac{CD}{Q^n}$$

where:

t = Liquid contact time, minutes
C = Constant
D = Filter depth, feet
Q = Hydraulic Loading rate, gpm/sq.ft.
n = Exponent characteristic of the filter media

Combining the time of contact with the first-order equation for BOD removal, in an adaptation of the Velz theory, Schulze derived the following formula:

$$\frac{L_e}{L_i} = e^{-KD/Q^n}$$

where:

L_e = BOD of unsettled filter effluent, mg/l
L_i = BOD of filter effluent
K = Treatability constant
n = Exponent characteristic of the filter media
D = Filter depth, feet

In 1965, Germain applied the Schulze formulation to a plastic media (Dowpac) filter as follows (12):

$$\frac{L_e}{L_o} = e^{-KD/Q^n}$$

where:

L_o = BOD of primary effluent (not including recirculation), mg/l
L_e = BOD remaining, mg/l
D = Depth of filter, feet
Q = Hydraulic load, gpm/sq.ft. (not including recirculation)
K = Treatability constant
n = Exponent characteristic of filter media

Germain found that K and n for Dowpac media treating domestic primary effluent were 0.088 and 0.5, respectively.

4.4.7 Eckenfelder Formula

In 1963, Eckenfelder modified the equations of Schulze to include the effect of changes in filter depth on the BOD removal per unit of depth. Eckenfelder proposed the following equations (23) (24):

$$\frac{L_e}{L_o} = \frac{1}{1 + \dfrac{CD^{(1-m)}}{\left(\dfrac{Q}{A}\right)^n}} \quad \text{and} \quad L_o = \frac{L_i + RL_e}{1 + R} \quad \text{where:}$$

L_i = Influent BOD (not including recirculation), mg/l
L_o = Influent BOD (including recirculation), mg/l
L_e = BOD of unsettled filter effluent, mg/l
R = Recirculation ratio
With A in acres, D in feet and Q in mgd:
$C = 2.5$; $1\text{-}m = 0.67$; $n = 0.5$

4.5 Applicability of Various Trickling Filter Design Formulas

The design engineer has available several formulas for trickling filter designs, and the decision to use one in preference to another is often difficult. The availability of several formulas often raises doubts concerning their validity in the mind of the design engineer.

An attempt has been made by Hanumanulu (25) to compare the actual performance of a 12-ft. deep stone media trickling filter with that predicted using NRC, Ten-States Standards, Velz, Eckenfelder, and Galler and Gotaas formulas. The filter was operated at a constant flow without recycle as well as with a 1:1 recirculation ratio. It was found that Velz, Ten-States Standards, and the NRC formulas predict filter efficiencies that are closer to observed values when operated without recycle, while the Eckenfelder and Galler and Gotaas formulas predict efficiencies closer to observed values for filters operated with recirculation.

Ordon (26) calculated the volume of filter media required to achieve specified BOD removals using the NRC, Eckenfelder, and Galler and Gotaas' formulas. The wastewater flow, BOD, and temperature were assumed as 1 mgd, 100 mg/l, and 20°C, respectively. The volume of filter calculated by the different formulas is shown in Table 4-2.

Inspection of Table 4-2 indicates characteristic trends which the designer should be aware of before using any of these formulas. In Table 4-2, when recirculation was zero, the filter volumes calculated from the NRC and Eckenfelder formulas were essentially the same, while the Galler and Gotaas formula gave volumes which were significantly different. However, when recirculation was considered, the NRC design volumes were generally quite conservative, while the volumes calculated by the Eckenfelder and Galler and Gotaas formulas were more nearly the same. In general, the NRC formulas would seem to apply when recirculation is not considered, when seasonal temperature differentials are minor, and when the wastewater load is highly variable and of high strength.

4.6 Laboratory and Pilot-Scale Treatability Studies

Trickling filters traditionally have been designed using one of the several formulas cited previously. The use of treatability studies for design of trickling filters has been hampered by the lack of suitable laboratory-scale testing methods, and has generally been restricted to the plastic-media filters, with pilot units being supplied by the manufacturers of plastic media on a rental basis. The pilot units available from the plastic media manufacturers require considerable manpower and funds to obtain the meaningful data needed for design purposes. Treatability studies for evaluation of stone-media filter design parameters are usually not performed.

However, it is interesting to note that advances are being made in the development of a practical laboratory-scale piloting facility for both stone and plastic media. Based on the concept of contact time as introduced by Schulze, the trickling filter process may be modeled by using an inclined plane to support biological growth (27).

Table 4-2

Trickling Filter Volumes for Various Organic Removals as Calculated by Different Design Formulas (All Volumes in Thousands of Cubic Feet)

Recirculation Ratio	50% BOD Removal			60% BOD Removal			70% BOD Removal			75% BOD Removal			80% BOD Removal			90% BOD Removal		
	NRC[1]	ECK[2]	G&G[3]	NRC	ECK	G&G	NRC	ECK	G&G	NRC	ECK	G&G	NRC	ECK	G&G	NRC	ECK	G&G
0	2.7	3.8	0.2	6	8.5	1.2	15	20	10	23	32	42	40	58	300	210	290	–
1	1.7	0.96	0.12	3.6	2.2	0.28	8.8	5	1.8	15	8	5	25	15	23	130	75	–
2	1.4	0.42	0.12	2.8	0.95	0.26	6.8	2.3	1.2	11	3.5	2.4	20	7	7.3	105	35	400
3	1.2	0.24	0.12	2.6	0.55	0.26	6.1	1.3	0.9	9.9	2	1.8	18	4	4.3	90	120	170
4	1.0	0.15	0.12	2.4	0.35	0.26	5.8	0.8	0.9	9.3	1.5	1.6	17	2.5	3.6	85	14	80
5	0.9	0.12	0.12	2.2	0.24	0.26	5.7	0.6	0.9	8.8	0.92	1.5	16	1.8	3.0	80	9	45
6	–	0.08	0.12	–	0.17	0.26	–	0.4	0.9	–	0.67	1.5	–	1.4	–	–	6	–

[1] NRC - National Research Council
[2] ECK - Eckenfelder
[3] G&G - Galler and Gotaas

Design Conditions
Filter Influent Flow = 1 mgd
Filter Influent BOD = 100 mg/l
Wastewater Temperature = 20°C

Wastewater is introduced at a variable rate to the top of a slimed inclined plane. The plane's inclination may be varied to change the contact time. As previously discussed, Schulze's formula relates the contact time to the depth and hydraulic loading, as well as to the physical characteristics of the filter media. BOD removal is then assumed to vary with the following first-order removal equation:

$$\frac{L_e}{L_i} = e^{-Kt}$$

where:

L_e = BOD of unsettled filter effluent, mg/l
L_i = BOD of filter influent, mg/l
K = Treatability constant
t = Contact time, minutes

The inclined plane method furnishes data on BOD removal, contact time, hydraulic loading, and recirculation ratios.

Since the basic purpose of either a laboratory or pilot-plant evaluation is to study variables that affect filter performance, any treatability studies should be of sufficient duration, and should consider the following variables as they affect the filter performance:

1. Applied BOD loading.
2. Hydraulic loading.
3. Recirculation.
4. Wastewater temperature.

The data thus obtained from treatability studies can be evaluated using the various trickling filter formulas previously discussed.

4.7 Trickling Filter Upgrading Techniques and Design Basis

Upgrading of trickling filter plants may be required because the plants are hydraulically and/or organically overloaded, because of the need for increased treatment efficiency, or both. Upgrading to relieve overloaded conditions and upgrading to improve removal efficiency to meet higher water quality standards will be covered in the following sections.

4.7.1 Upgrading to Relieve Organic and Hydraulic Overloading

Trickling filter plants may be upgraded to relieve hydraulic and/or organic overloading by any one of the following three general procedures:

1. Upgrading an existing single-stage filter to adequately handle an increased load, either organic or hydraulic.

2. Upgrading a single-stage trickling filter to a two-stage biological system
3. Upgrading an existing two-stage trickling filter to a multiple-stage biological system.

There are several factors that should be considered prior to upgrading a trickling filter plant. Since upgrading varies from plant to plant, only general observations can be made. Items to be considered are:

1. Check the hydraulic capacity of the trickling filter distributor arm to determine the recommended operating range.

2. Investigate the ventilation in all pipes, channels, and drains. As previously discussed, not more than 50 percent of any conduit's cross-section should be submerged under average hydraulic design loading.

3. Decide whether to use direct recirculation after the filter or recirculation of the clarified effluent. One study has indicated that direct recirculation of filter effluent is as effective as recycling clarified effluent (28).

4. Evaluate the capability of the secondary clarifier to determine if additional capacity is required and if the sludge-collection mechanism is performing correctly.

5. Check and evaluate the capacity of the sludge-handling facilities. Upgrading secondary treatment facilities usually results in an increased sludge production.

The following examples are illustrative in nature and are not based on actual performance data unless specified in the text.

4.7.1.1 Upgrading a Single-Stage Trickling Filter-Conversion From Low-Rate to High-Rate (Example A)

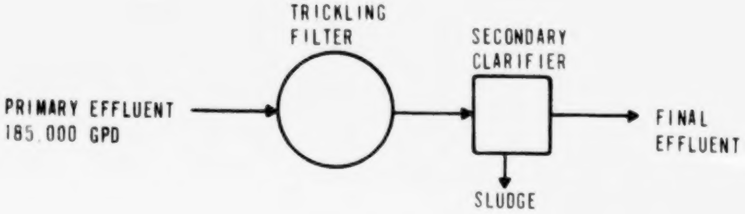
Upgrading of a hydraulically overloaded low-rate trickling filter can be accomplished by converting it to a high-rate trickling filter through recirculation. This upgrading procedure has been used successfully in the following areas: Edgerton, Wisconsin; Flandreau, South Dakota; Pueblo, Colorado; and Coeur d'Alene, Idaho (29).

Example A will help to illustrate the design considerations involved in upgrading a low-rate filter to a high-rate filter. A flow diagram for the overloaded plant appears in Figure 4-8. Table 4-3 contains a comparison of the original design values of the low rate filter before it was overloaded, as well as the operating data from the overloaded plant before upgrading. The flow increased from 185,000 gpd to 370,000 gpd, while the BOD and suspended solids in the effluent increased from 30 mg/l and 23 mg/l, respectively, to 44 mg/l and 36 mg/l.

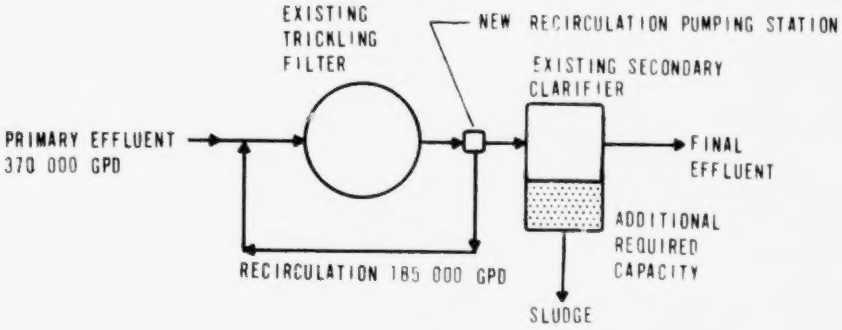
FIGURE 4-8

UPGRADING A LOW-RATE TRICKLING FILTER TO A HIGH-RATE TRICKLING FILTER

EXAMPLE A


TRICKLING FILTER
SECONDARY CLARIFIER
PRIMARY EFFLUENT 185,000 GPD
FINAL EFFLUENT
SLUDGE

TREATMENT SYSTEM BEFORE UPGRADING
LOW-RATE TRICKLING FILTER


EXISTING TRICKLING FILTER
NEW RECIRCULATION PUMPING STATION
EXISTING SECONDARY CLARIFIER
PRIMARY EFFLUENT 370 000 GPD
FINAL EFFLUENT
ADDITIONAL REQUIRED CAPACITY
RECIRCULATION 185 000 GPD
SLUDGE

TREATMENT SYSTEM AFTER UPGRADING
HIGH-RATE TRICKLING FILTER

Table 4-3

Operational and Design Data for Example A

Description	Original Design Before Overloading	Overloaded Operating Condition	Upgraded Design Calculations
Flow - gpd	185,000	370,000	370,000
Influent BOD, mg/l	230	210	210
Influent SS, mg/l	210	200	200
Primary Clarifier			
Overflow Rate, gpd/sq.ft.	700	1,400	700
Percent BOD Removal	32	23	32
Percent SS Removal	55	48	55
Trickling Filter			
Depth, feet	6	6	6
Hydraulic Loading, mgad[1]	4.6	9.2	13.8
Organic Loading, lbs. BOD/1,000 cu.ft./day[1]	23	48	47
Recirculation Ratio	0	0	0.5
Secondary Clarifier			
Overflow Rate, gpd/sq.ft.	800	1,600	800
Secondary Treatment			
Percent BOD Removal	81	73	79
Percent SS Removal	75	65	75
Overall Plant Performance			
Percent BOD Removal	87	79	86
Percent SS Removal	89	82	89
Effluent BOD, mg/l	30	44	30
Effluent SS, mg/l	23	36	22

[1] Includes recirculation.

To upgrade the plant to its previous performance, it was decided to renovate the plant so that it could treat the flow of 370,000 gpd as a high-rate filter. The initial step was to evaluate the quantity of recycled flow to be returned ahead of the filter. This can be done by using one of the trickling filter design formulas presented in the previous section. The results of the upgrading calculations are presented in Table 4-3, and the upgraded flow diagram is presented in Figure 4-8.

To implement this upgrading, several factors were investigated. The hydraulic capacity of the existing distributor arm was found to be deficient and therefore replaced. The hydraulic head available to the filter was found to be limiting; therefore, the new distributor arm was motorized. The existing filter media and underdrains were found to be in good condition. The hydraulic capacity of the drains was evaluated and found to be sufficient. A recirculation pumping station was constructed with variable-speed pumping capacity regulated with flow-proportioning pump controls. The final clarification capacity was increased to accommodate the larger flows. In addition, the primary clarification capacity would also have to be increased, but this cost is not considered in this unit operations section.

The capital costs associated with this upgrading were estimated at $96,000 ($519 per 1,000 gpd of incremental upgraded capacity) and were allocated as follows:

Trickling Filter Modification	$51,000
Recirculation Facilities	15,000
Secondary Clarifier Expansion	30,000
TOTAL	$96,000[1]

4.7.1.2 Upgrading a Single-Stage Trickling Filter - Conversion from High-Rate to Completely-Mixed Activated Sludge (Example B)

In 1965, the Ontario Water Resources Commission set 15 mg/l of suspended solids and BOD as the objectives for secondary treatment plant effluents. Such an effluent quality could not be achieved with an existing high-rate trickling filter plant at Gravenhurst, Ontario (30).

To upgrade the high-rate filter, the plant was converted to completely-mixed activated sludge. The filter was converted to an aeration tank 40 feet in diameter by removing the media and raising the concrete sidewalls by seven feet to a total height of 12 feet. A 10-hp mechanical aerator was installed. The duo-clarifier (combination primary and secondary clarifier) was converted to a 40-foot diameter secondary clarifier, and a new 35-foot diameter primary clarifier was constructed. A 100-percent sludge recycle capacity was provided.

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[1] These costs are based on ENR index of 1500 and contain no contingency for engineering design, bonding, and construction supervision.

The previously described upgrading technique resulted in the following measured improvements:

Parameter	Before Upgrading	After Upgrading
Dry weather design flow, gpd	300,000	375,000
Influent organic load, lbs. BOD/day	360	540
Effluent BOD, mg/l	>20	15-20

The capital costs for this upgrading were estimated at $70,000 and were allocated as follows:

Tank modification	$55,000
Secondary clarifier modification	15,000
TOTAL	$70,000[1]

These costs do not include upgrading of any other unit treatment processes, e.g., primary clarification.

4.7.1.3 Upgrading a Single-Stage Trickling Filter to a Two-Stage Biological System - Conversion From a Single-Stage to a Two-Stage Filtration System (Example C)

Organically overloaded low, intermediate, and high-rate trickling filters may be upgraded by converting them to two-stage filtration systems, utilizing rock media for both stages. Example C depicts such an upgrading, and it will illustrate the major considerations to be evaluated.

A flow diagram for Example C before upgrading is shown in Figure 4-9. A summary of the operating data is presented in Table 4-4. Treating a flow of 6 mgd, the intermediate-rate filter produced a final effluent with BOD and SS concentrations of 99 mg/l and 85 mg/l, respectively. To improve the organically overloaded conditions, it was decided to design a high-rate filter and intermediate clarifier to operate ahead of the existing, intermediate-rate filter. The appropriate recirculation ratio and filter volume were calculated using one of the design formulas previously discussed. The results of these

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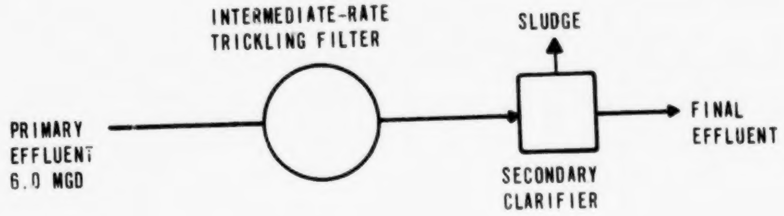
[1] These costs are based on ENR index of 1500 and contain no contingency for engineering design, bonding, and construction supervision.

FIGURE 4-9

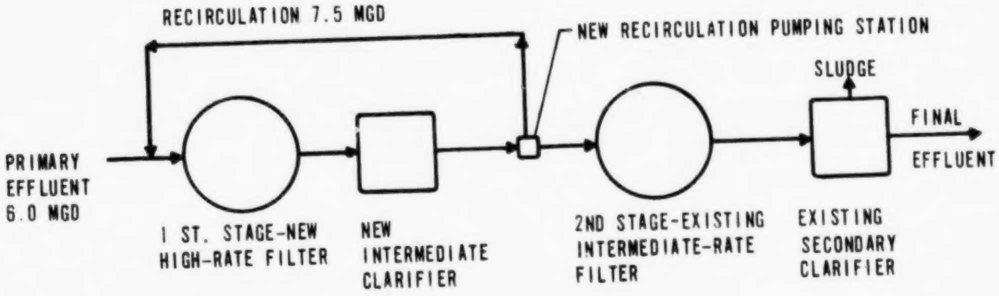
UPGRADING A SINGLE-STAGE TRICKLING FILTER TO A TWO-STAGE FILTRATION SYSTEM

EXAMPLE C


INTERMEDIATE-RATE
TRICKLING FILTER
SLUDGE
PRIMARY
EFFLUENT
6.0 MGD
FINAL
EFFLUENT
SECONDARY
CLARIFIER

TREATMENT SYSTEM BEFORE UPGRADING

SINGLE-STAGE INTERMEDIATE-RATE TRICKLING FILTER


RECIRCULATION 7.5 MGD
NEW RECIRCULATION PUMPING STATION
SLUDGE
PRIMARY
EFFLUENT
6.0 MGD
FINAL
EFFLUENT
1 ST. STAGE-NEW
HIGH-RATE FILTER
NEW
INTERMEDIATE
CLARIFIER
2ND STAGE-EXISTING
INTERMEDIATE-RATE
FILTER
EXISTING
SECONDARY
CLARIFIER

TREATMENT SYSTEM AFTER UPGRADING

TWO-STAGE TRICKLING FILTRATION SYSTEM

(BOTH STAGES UTILIZE ROCK MEDIA)

Table 4-4

Operational and Design Data for Example C

Description	Overloaded Operating Condition	Upgraded Design Calculations
Flow - mgd	6	6
Influent BOD, mg/l	355	355
Influent SS mg/l	340	340
Primary Clarifier		
Overflow Rate, gpd/sq.ft.	750	750
Percent BOD Removal	38	38
Percent SS Removal	60	60
Trickling Filter - 1st Stage		
Depth, feet	–	4
Hydraulic Loading, mgad[1]	–	19
Organic Loading, lbs. BOD/1,000 cu.ft./day[1]	–	110
Recirculation Ratio	–	1.25
Intermediate Clarifier		
Overflow Rate, gpd/sq.ft.[1]	–	1,000
Percent BOD Removal - 1st Stage	–	81.7
Percent SS Removal - 1st Stage	–	70.0
Trickling Filter - 2nd Stage		
Depth, feet	7	7
Hydraulic Loading, mgad[1]	8.3	8.3
Organic Loading, lbs. BOD/1,000 cu.ft./day[1]	50	10
Recirculation Ratio	0	0
Final Clarifier		
Overflow Rate, gpd/sq.ft.	800	800
Percent BOD Removal - 2nd Stage	55	50
Percent SS Removal - 2nd Stage	38	63
Overall Plant Performance		
Percent BOD Removal	72	94
Percent SS Removal	75	96
Effluent BOD, mg/l	99	20
Effluent SS, mg/l	85	15

[1] Includes recirculation.

calculations are presented in Table 4-4, and the upgraded flow diagram is shown in Figure 4-9. The upgraded effluent is expected to contain 20 mg/l of BOD and 15 mg/l of SS.

This type of upgrading, in which a complete set of units is added, is far less complicated than a renovation of existing tankage. The most important consideration in this type of upgrading is that sufficient hydraulic head be available to operate the individual unit processes properly. In Example C, the major capital costs include providing new high-rate filters, intermediate clarifiers, a recirculation pumping station regulated with flow-proportioning pump controls, and the appropriate piping. The capital costs associated with this upgrading were estimated to be $1,500,000 and were allocated as follows:

Trickling Filter Additions	$1,000,000
Recirculation Facilities	100,000
Intermediate Clarification	400,000
TOTAL	$1,500,000[1]

4.7.1.4 Upgrading a Single-Stage Trickling Filter to a Two-Stage Biological System - Conversion of a Single-Stage Filter to a Filtration/Activated Sludge System (Example D)

If the hydraulic and organic loads to a high-rate filter are such that it would not produce a high degree of BOD removal, it is possible to upgrade the facility by the addition of an activated sludge unit immediately downstream from the existing filter. In this situation, the existing overloaded trickling filter acts as a roughing filter, and the subsequent activated sludge unit provides the treatment capacity needed to obtain the desired BOD removal.

Example D illustrates the major considerations in this type of upgrading. The flow diagram of the overloaded plant appears in Figure 4-10, and operating data for the overloaded period are presented in Table 4-5. The existing plant was upgraded by the addition of a completely-mixed activated sludge system The calculations for the upgrading are summarized in Table 4-5, and the upgraded flow diagram is shown in Figure 4-10. Implementation of this upgrading technique would reduce the effluent BOD an estimated 150 mg/l, from 220 mg/l to 70 mg/l. Details concerning the design of an activated sludge system treating an effluent from a single-stage biological treatment process are presented in a later section.

Construction costs for Example D include the costs for a completely-mixed aeration basin, floating mechanical aerators, and an activated sludge recirculation pumping station. In

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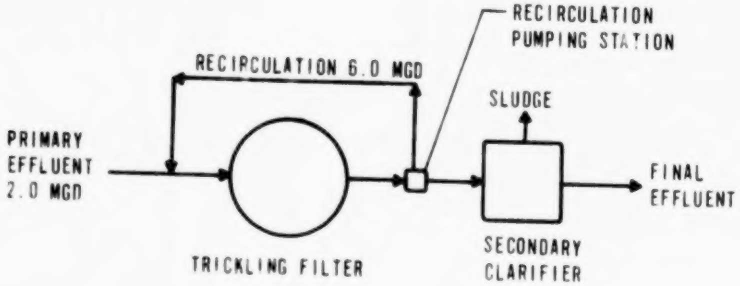
[1] These costs are based on an ENR index of 1500 and contain no allowance for engineering design, bonding, and construction supervision.

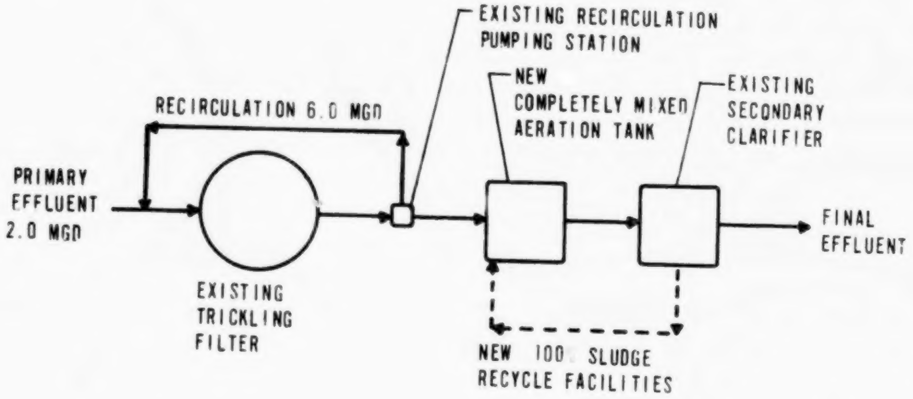
FIGURE 4-10

UPGRADING A HIGH-RATE TRICKLING FILTER TO A TWO-STAGE FILTRATION/ACTIVATED SLUDGE SYSTEM

EXAMPLE D


RECIRCULATION PUMPING STATION
RECIRCULATION 6.0 MGD
SLUDGE
PRIMARY EFFLUENT 2.0 MGD
FINAL EFFLUENT
TRICKLING FILTER
SECONDARY CLARIFIER

TREATMENT SYSTEM BEFORE UPGRADING
HIGH-RATE TRICKLING FILTER


EXISTING RECIRCULATION PUMPING STATION
NEW COMPLETELY MIXED AERATION TANK
EXISTING SECONDARY CLARIFIER
RECIRCULATION 6.0 MGD
PRIMARY EFFLUENT 2.0 MGD
FINAL EFFLUENT
EXISTING TRICKLING FILTER
NEW 100% SLUDGE RECYCLE FACILITIES

TREATMENT SYSTEM AFTER UPGRADING
FILTRATION/ACTIVATED SLUDGE SYSTEM

Table 4-5

Operational and Design Data for Example D

Description	Overloaded Operating Conditions	Upgraded Design Calculations
Flow, mgd	2	2
Influent BOD, mg/l	550	550
Influent SS, mg/l	400	400
Primary Clarifier		
Overflow Rate, gpd/sq.ft.	700	700
Percent BOD Removal	20	20
Percent SS Removal	40	40
Trickling Filter		
Depth, feet	6	6
Hydraulic Loading, mgad[1]	50	50
Organic Loading, lbs. BOD/1,000 cu.ft./day[1]	260	260
Recirculation Ratio	3.0	3.0
Percent BOD Removal as a Roughing Unit	–	50
Completely-Mixed Aeration Tank		
Detention Time Based on Average Flow, hours[1]	–	3.0
Sludge Recycle Capacity, percent of design flow	–	100
Volumetric Loading, lbs. BOD/day/1,000 cu.ft. of Aeration Tank Volume	–	50
Secondary Clarifier		
Overflow Rate, gpd/sq.ft.	700	700
Percent BOD Removal[2]	50	84
Percent SS Removal[2]	40	67
Overall Plant Performance		
Percent BOD Removal	60	87
Percent SS Removal	50	80
Effluent BOD, mg/l	220	70
Effluent SS, mg/l	144	80

[1] Includes recirculation

[2] In secondary units including the roughing filter.

addition, the secondary clarifier was modified to use a suction-type sludge removal mechanism. The cost for this upgrading has been estimated at $320,000, and is broken down as follows:

Aeration Tank	$190,000
Sludge Recirculation	70,000
Clarifier Modifications	60,000
TOTAL	$320,000[1]

4.7.1.5 Upgrading a Single-Stage Trickling Filter to a Two-Stage Biological System - Addition of a Super-Rate Roughing Filter to a Single-Stage Trickling Filter (Example E)

An organically overloaded high-rate trickling filter may be upgraded by placing a synthetic media super-rate filter immediately upstream to act as a roughing unit. Example E is presented to illustrate the engineering considerations which must be evaluated.

Figure 4-11 contains flow diagrams of the secondary treatment system before and after upgrading. Table 4-6 contains operational data from the overloaded plant. The roughing filter was sized using the Schulze formula presented previously. Design data for the roughing filter is also summarized in Table 4-6. By removing 30 percent of the applied BOD in the roughing filter (not including recirculation), it was possible to reduce the recirculation ratio from 3 to 2 on the existing high-rate filter.

The construction costs include the roughing filter, a recirculation pumping station with flow-proportioning controls, and all appropriate piping. This upgrading was estimated to cost $215,000 and is allocated as follows:

Roughing Filter	$190,000
Recirculation Facilities	25,000
TOTAL	$215,000[1]

4.7.1.6 Upgrading an Existing Two-Stage Trickling Filter to a Multiple-Stage Biological System

Fewer options are available for upgrading a hydraulically or organically overloaded two-stage filter than for upgrading a single-stage filter. In general, there are three options available to the engineer faced with upgrading an overloaded two-stage filter:

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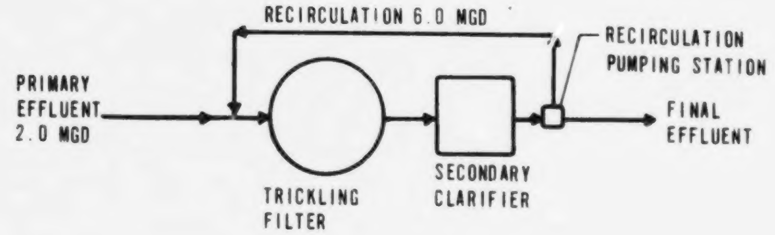
[1] These costs are based on an ENR index of 1500 and contain no allowance for engineering design, bonding, and construction supervision.

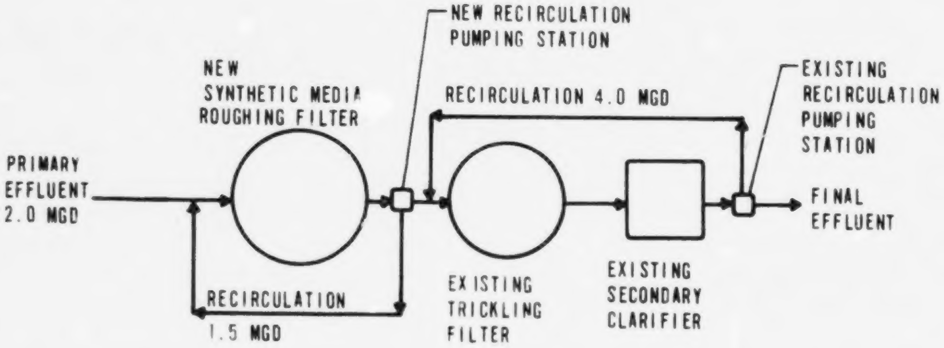
FIGURE 4-11

UPGRADING A HIGH-RATE TRICKLING FILTER USING A SUPER-RATE TRICKLING FILTER AS A ROUGHING UNIT

EXAMPLE E


RECIRCULATION 6.0 MGD
RECIRCULATION PUMPING STATION
PRIMARY EFFLUENT 2.0 MGD
FINAL EFFLUENT
SECONDARY CLARIFIER
TRICKLING FILTER

TREATMENT SYSTEM BEFORE UPGRADING
HIGH-RATE TRICKLING FILTER


NEW RECIRCULATION PUMPING STATION
NEW SYNTHETIC MEDIA ROUGHING FILTER
RECIRCULATION 4.0 MGD
EXISTING RECIRCULATION PUMPING STATION
PRIMARY EFFLUENT 2.0 MGD
FINAL EFFLUENT
RECIRCULATION 1.5 MGD
EXISTING TRICKLING FILTER
EXISTING SECONDARY CLARIFIER

TREATMENT SYSTEM AFTER UPGRADING
ROUGHING FILTER PRECEEDING EXISTING HIGH-RATE FILTER

Table 4-6

Operational and Design Data for Example E

Description	Overloaded Operating Conditions	Upgraded Design Calculations
Flow, mgd	2	2
Influent BOD, mg/l	550	550
Influent SS, mg/l	400	400
Primary Clarifier		
Overflow Rate, gpd/sq.ft.	700	700
Percent BOD Removal	20	20
Percent SS Removal	40	40
Roughing Filter		
Depth, feet	–	11
Hydraulic Loading, mgad[1]	–	62.6
Organic Loading, lbs. BOD/1,000 cu.ft./day[1]	–	430
Recirculation Ratio	–	0.75
Percent BOD Removal	–	30
Trickling Filter		
Depth, feet	6	6
Hydraulic Loading, mgad[1]	50	38
Organic Loading, lbs. BOD/1,000 cu.ft./day[1]	260	147
Recirculation Ratio	3.0	2.0
Secondary Clarifier		
Overflow Rate, gpd/sq.ft.[1]	700	520
Percent BOD Removal[2]	50	84
Percent SS Removal[2]	40	67
Overall Plant Performance		
Percent BOD Removal	60	87
Percent SS Removal	50	80
Effluent BOD, mg/l	220	70
Effluent SS, mg/l	144	80

[1] Includes recirculation

[2] In secondary units including roughing unit

1. Construction of a roughing filter preceding the existing system.
2. Construction of an activated sludge system following the existing system.
3. Construction of a separate parallel biological treatment system.

A detailed discussion will not be presented here, since most of the engineering considerations pertaining to these three options have been examined in previous sections.

4.7.2 Upgrading to Increase Organic Removal Efficiency

Upgrading techniques previously discussed relate to the ability of existing facilities to handle increased hydraulic or organic loads by providing modifications to meet existing effluent standards. However, there may be a need to meet higher effluent standards even though the existing facilities are not hydraulically or organically overloaded. Table 4-7 contains suggested alternatives for improving effluent quality under these conditions. The main purpose of the table is to present various alternatives and to suggest a range of anticipated improvement in performance for each alternative.

It should be emphasized that, in cases where unit processes are added to existing facilities, the improvement in overall organic removal will be a direct function of the BOD removal achieved in the "add-on" unit process, e.g., a polishing lagoon. However, where unit processes precede existing units, e.g., the use of a roughing filter, the overall BOD removal may not be increased in direct proportion to the amount achieved in the "add-on" process.

A detailed discussion on polishing lagoons, microstrainers, filters, activated carbon, and clarifier modifications appears in subsequent chapters. The applicability of alternatives to individual cases should be evaluated in detail prior to the implementation of a particular upgrading procedure.

Table 4-7

Upgrading Techniques for Improvement of Trickling Filter Plant Efficiency

Addition Preceding Existing Unit	Modification to Existing Unit	Addition Following Existing Unit	Incremental BOD Removal Across the Added or Modified Process percent
	1. Low-Rate Trickling Filter		
	Add Recirculation during low-flow periods		0-10
	2. High-Rate Trickling Filter		
	Increase Recirculation		0-10
	3. Two-Stage Trickling Filter		
	--[1]		--
Roughing Trickling Filter (Rock or Synthetic Media)			20-40
Chemical Addition To Primary Clarifier			30-50
		2nd Stage Activated Sludge[2]	30-70
		Polishing Lagoon	30-60
		Multi-media Filters	50-80
		Microstraining	30-80
		Activated Carbon	60-80

[1] Generally not amenable to modifications for increasing treatment efficiency.

[2] A consideration if year-round nitrification is required.

4.8 References

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CHAPTER 5

TECHNIQUES FOR UPGRADING ACTIVATED SLUDGE PLANTS

5.1 General

The conventional activated sludge process as originally developed has undergone significant changes, primarily due to a better understanding of the theory involved and to the experience accumulated over the years in successful operation of the process. Today, it remains the most versatile and efficient of the available biological treatment processes.

Historically, the activated sludge process has been used in larger cities, where the ratio of river assimilative capacity to waste load is small. In the past decade, there has been a trend toward its use by smaller communities to meet the more stringent requirements stipulated by regulatory agencies.

Existing overloaded conventional activated sludge plants pose a problem to the maintenance of established water quality standards. Various modifications of the conventional process developed over the years permit reduced detention time in the aeration tanks; the applicability of these process modifications in the efficient upgrading of existing plants will be examined and discussed.

5.2 Activated Sludge Processes

Basically, the activated sludge process uses microorganisms in suspension to oxidize soluble and colloidal organics to CO_2 and H_2O in the presence of molecular oxygen. During the oxidation process, a portion of the organic material is synthesized into new cells. A part of the synthesized cells then undergoes auto-oxidation in the aeration tanks, the remainder forming excess sludge. Oxygen is required in the process to support the oxidation and synthesis reactions. In order to operate the process on a continuous basis, the solids generated must be separated in a clarifier for recycle to the aeration tank, with the excess sludge from the clarifiers being withdrawn for further handling and disposal.

5.2.1 Conventional Activated Sludge

A schematic of the conventional activated sludge process is shown in Figure 5-1. The wastewater is commonly aerated for a period of 6 to 8 hours (based on the average design flow) in the presence of a portion of the secondary sludge (1). The rate of sludge return expressed as a percentage of the average wastewater design flow is normally about 25 percent, with minimum and maximum rates of 15 and 75 percent. The plug flow mixing configuration is used to condition the biological organisms for improved clarification. This is accomplished in rectangular tanks, designed so that the total tank length is generally 5 to 50 times the width. Operational data from various conventional activated sludge plants are summarized in Table 5-1.

Table 5-1

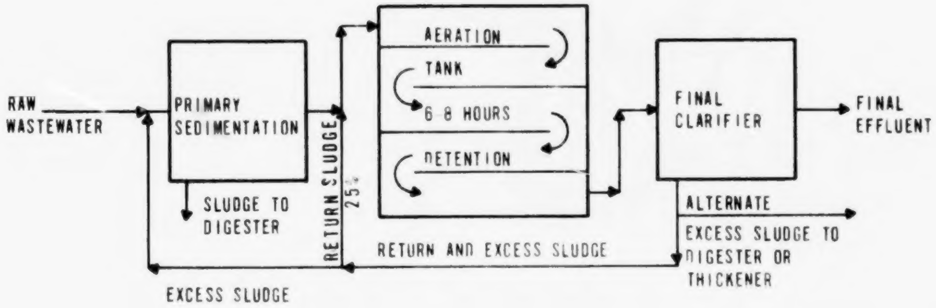
Operational Data from Various Conventional Activated Sludge Plants

Plant Location	Influent Flow mgd	Sludge Recycle percent	BOD Influent mg/l	BOD Effluent mg/l	Aeration Tank MLSS mg/l	Organic Loading lbs. BOD/day / lb. MLSS	Volumetric Loading lbs. BOD/day / 1,000 cu.ft.	Aeration Detention Time[1] hours	Air Supplied per lb. of BOD Removed cu.ft./lb	Secondary BOD Removal Efficiency percent	Reference
Michigan	5.0	32.0	182	19	1,844	0.34	39	6.99	770	89.6	2
Illinois	288.0	47.9	129	11	1,930	0.17	21	8.7	876	91.5	2
Ohio	86.9	25.0	91	12	2,180	0.12	17	7.73	1,600	86.9	2
Indiana	14.9	30.0	161	14	2,420	0.16	24	10.0	733	91.0	2
Maryland	3.9	32.0	254	32	1,808	0.39	44	8.8	500	86.8	2
Michigan	8.0	15.5	118	6	2,801	0.15	26	6.7	690	94.9	2
Wisconsin	7.6	51.6	157	36	1,094	0.38	26	9.1	690	77.1	2
Indiana	3.9	30.8	134	14	2,625	0.21	35	5.7	886	89.7	2
Indiana	5.5	28.5	113	6	1,680	0.19	20	8.2	435	94.7	2
Maryland	8.0	26.0	155	10	2,040	0.23	29	7.7	1,260	93.5	3
Maryland	7.7	25.0	148	15	2,240	0.20	25	8.2	1,900	89.9	3

[1] Excluding sludge recycle

FIGURE 5-1

CONVENTIONAL ACTIVATED SLUDGE FLOW DIAGRAM


RAW WASTEWATER
PRIMARY SEDIMENTATION
SLUDGE TO DIGESTER
RETURN SLUDGE 25%
AERATION
TANK
6-8 HOURS
DETENTION
FINAL CLARIFIER
FINAL EFFLUENT
ALTERNATE
EXCESS SLUDGE TO DIGESTER OR THICKENER
RETURN AND EXCESS SLUDGE
EXCESS SLUDGE

The following factors have been cited (1) as limitations in the design and use of the conventional activated sludge process:

1. BOD loadings are limited to about 35 lbs./1,000 cu.ft./day.
2. A high initial oxygen demand is experienced in the head end of the aeration tank.
3. The final clarifier is subjected to high solids loadings.
4. It is necessary to increase sludge recirculation proportionately with increasing BOD loadings.
5. Detention times are in the range of 6 to 8 hours.
6. There is a lack of operational stability with variations in hydraulic and organic loadings.

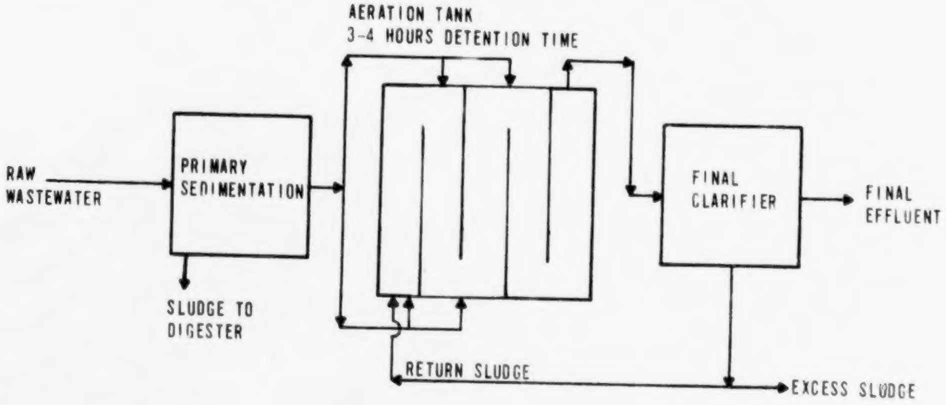
Some of these limitations have stimulated the development and use of various activated sludge modifications, such as step aeration, contact stabilization, completely-mixed, two-stage activated sludge, and the use of oxygen aeration instead of air as a source of dissolved oxygen. These modifications are discussed in subsequent sections.

5.2.2 Step Aeration

The step aeration modification is illustrated in Figure 5-2. Unlike the conventional flow pattern, the influent wastewater is introduced at various points along the length of the aeration tank. The Ten-States Standards recommends a sludge return rate (based on the average wastewater flow) of 50 percent, with minimum and maximum rates of 20 and 75 percent. In actual cases, this rate has been found to be as high as 100 percent. Splitting up the influent flow to the aeration tank reduces the initial oxygen demand usually experienced in the conventional process, and distributes the organic loading more uniformly over the length of the aeration tank. This appears to afford a more efficient utilization of the biological population of the tank, which explains the fact that organic loadings up to 50 lbs. BOD/1,000 cu.ft./day have been treated. Operational data from various step aeration processes are summarized in Table 5-2.

FIGURE 5-2

STEP AERATION FLOW DIAGRAM


AERATION TANK
3-4 HOURS DETENTION TIME
RAW WASTEWATER
PRIMARY SEDIMENTATION
FINAL CLARIFIER
FINAL EFFLUENT
SLUDGE TO DIGESTER
RETURN SLUDGE
EXCESS SLUDGE

There is some question regarding the air requirements for this modification. Generally, the step aeration process will not utilize any more air than a conventional system treating comparable flows. In some cases, such as in New York City (4) where the sludge recycle is approximately 25 percent, the air required is about half that normally used for a conventional process.

Table 5-2

Operational Data from Various Step Aeration Plants

Plant Location	Influent Flow mgd	Sludge Recycle percent	BOD Influent mg/l	BOD Effluent mg/l	Aeration Tank MLSS mg/l	Organic Loading lbs. BOD/day / lb. MLSS	Volumetric Loading lbs. BOD/day / 1,000 cu ft	Air Supplied cu ft/gal	Air Supplied per lb. of BOD Removed cu ft/lb	Aeration Detention Time[1] hours	Secondary BOD Removal Efficiency percent	Reference
New York	110	24	74	12	1,170	0.49	36	–	910	3.1	83.8	4
New York	20.7	49	137	3	3,520	0.10	23	–	910	8.4	94.2	4
New York	92	35	100	8	1,110	0.42	30	–	933	4.9	92.2	4
New York	50	28	120	6	3,300	0.31	71	0.43	–	2.5	94.0	4
New York	95	28	115	16	3,300	0.28	58	0.54	–	2.9	86.0	4
New York	31	28	100	12	4,400	0.13	37	0.59	–	4.2	90.0	4
Maryland	16.9	24	140	11	2,120	0.54	58	–	–	3.8	92.3	3
Washington, D.C.	0.125	65	82	18	6,400	0.08	32	1.5-2.0	–	3.8	84.0	5
Washington, D.C.	0.125	50	110	32	2,050	0.37	48	1.5-2.0	–	5.0	71.0	5
Indiana	12.8	92	124	15	2,900	0.19	33	–	1,240	5.3	88.8	4
Indiana	12.4	77.3	134	14	2,600	0.17	29	–	1,090	7.0	88.8	4
Indiana	19.3	49.7	139	17	2,750	0.22	41	–	1,080	5.0	87.8	4
Indiana	34.3	31.7	131	18	3,360	0.22	45	–	911	4.3	86.3	4

[1] Excluding sludge recycle

This decrease in air requirements is attributed to its more effective utilization. In step aeration systems which utilize higher sludge recycle, the air requirements approach those of the conventional system. A significant design consideration in these latter step aeration systems is that since the detention times are lower than for the conventional system, the air supply system and diffusion equipment must be modified to supply approximately the conventional volume of air to a tank approximately one-half the conventional size.

In the conventional process, the mixed liquor concentration is intended to be relatively constant throughout the aeration tank, while in the step aeration process the concentration decreases as the return sludge becomes further diluted with the influent flow This principle is illustrated in Figure 5-3 (6). A lower solids loading may, in some cases, improve clarifier performance.

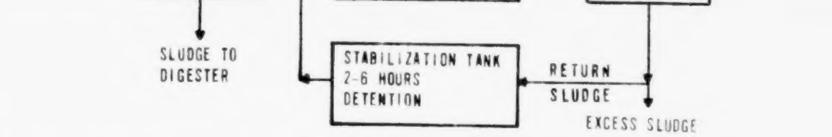
5.2.3 Contact Stabilization

The principles involved in the contact stabilization modification were initially demonstrated in the upgrading of an existing hydraulically overloaded conventional plant in Austin, Texas. The design capacity was upgraded from 6 mgd to 15 mgd using a contact stabilization flow pattern (7).

Laboratory studies and field work have demonstrated that wastewater BOD in the colloidal or insoluble state is rapidly removed from wastewater in a relatively short contact time by the combined physical processes of biological flocculation, adsorption, and enzyme-complexing. This offers the possibility of substantial reduction in plant volume for wastewaters largely in these forms. In the contact stabilization process, after the biological sludge is separated from the wastewater in the clarifier, the concentrated sludge is further aerated in another aeration tank (called the stabilization tank). Here, the flocculated and adsorbed BOD is stabilized (Figure 5-4). In addition to a shorter total contact time, the contact stabilization modification has the advantage of being able to handle greater shock and toxic loadings because of the biological buffering capacity of the stabilization tank, and the fact that at any given time the majority of the activated sludge is isolated from the main stream of the plant flow.

FIGURE 5-4

CONTACT STABILIZATION FLOW DIAGRAM

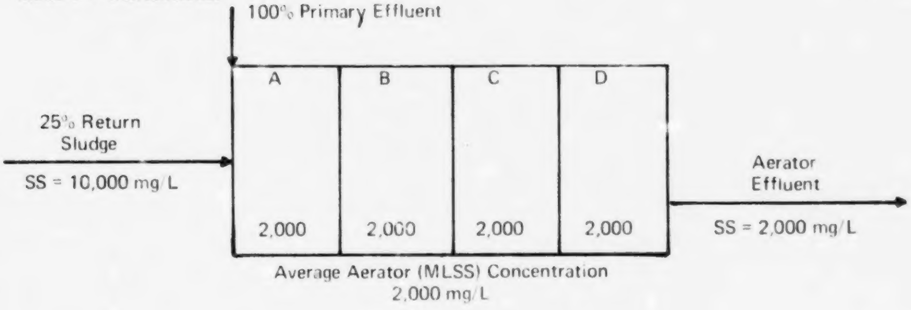


Operational data from four contact stabilization processes are summarized in Table 5-3.

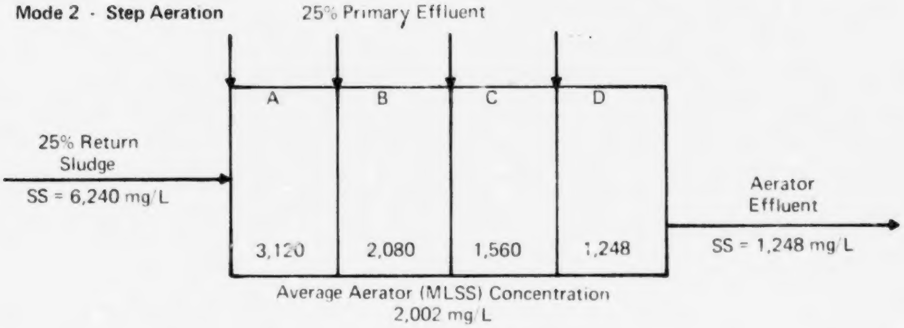
FIGURE 5-3

COMPARISON OF SOLIDS LOADING ON THE FINAL CLARIFIER FOR CONVENTIONAL AND STEP AERATION SYSTEMS

Mode 1 - Conventional


100% Primary Effluent
A
B
C
D
25% Return Sludge
SS = 10,000 mg/L
2,000
2,000
2,000
2,000
Aerator Effluent
SS = 2,000 mg/L
Average Aerator (MLSS) Concentration
2,000 mg/L

Mode 2 - Step Aeration


25% Primary Effluent
A
B
C
D
25% Return Sludge
SS = 6,240 mg/L
3,120
2,080
1,560
1,248
Aerator Effluent
SS = 1,248 mg/L
Average Aerator (MLSS) Concentration
2,002 mg/L

Example assumes negligible suspended solids in the primary effluent and final effluent

Table 5-3

Operational Data from Various Contact Stabilization Plants

Plant Location	Influent Flow mgd	Sludge Recycle percent	BOD Influent mg/l	BOD Effluent mg/l	Contact Tank MLSS mg/l	Contact Tank Aeration Time[1] minutes	Stabilization Tank MLSS mg/l	Stabilization Tank Aeration Time[2] minutes	Organic Loading lbs. BOD/day / lb. MLSS	Volumetric Loading lbs. BOD/day / 1,000 cu.ft.	Secondary BOD Removal Efficiency percent	Reference
Texas	8.3	37.4	200	24	2,530	28	7,700	148	0.25	108	88.0	7
Texas	7.2	41.3	228	10	3,930	33	1,000	172	0.17	99	95.6	7
Texas	7.8	32.7	195	14	2,620	30	6,850	156	0.25	97	92.8	7
Texas	8.0	33.3	295	28	2,440	29	7,500	142	0.35	148	90.5	7
Texas	7.4	34.6	262	12	3,210	32	7,320	168	0.23	94	95.4	7
Texas	7.4	35.1	208	10	3,020	32	8,580	168	0.43	156	95.2	7
Texas	7.1	50.2	333	25	2,760	33	6,300	170	0.49	196	92.5	7
Texas	7.0	51.4	320	24	2,290	34	7,550	175	0.35	148	92.5	7
Texas	5.1	71.1	216	16	2,540	46	6,700	240	0.17	69	92.6	7
New Jersey	2.5	70.0	312	30	4,000	100	6,700	144	0.32	104	90.4	8
Maryland	8.0	50.0	200	8	1,620	78	4,200	78	0.38	–	96.0	3

[1] Based on influent flow excluding sludge recycle.
[2] Based on sludge recycle flow.

The time required for stabilization is a function of contact time, temperature, and strength of the waste. An increase in contact time normally reduces the stabilization time requirements (9). An important design consideration is the need for adequate stabilization to ensure satisfactory secondary clarifier performance. The relative detention times and air requirements of the contact and stabilization tanks have been reported by Lesperance (9) and are summarized in Table 5-4.

Table 5-4

Comparison of Contact Stabilization Detention Times and Air Requirements

Detention Time hours	Contact Tank based on forward plus recycle flow	Stabilization Tank based on sludge recycle
Wastewater containing mostly insoluble BOD (domestic wastewater), minimum	0.5	2.0
Wastewater containing mostly soluble BOD, minimum	2	3
Most commonly used	0.5 to 1.0	2 to 6

Air Requirements percent of total	Contact Tank percent	Stabilization Tank percent
Wastewater containing mostly insoluble BOD	40	60
Wastewater containing mostly soluble BOD	60	40

Most of the benefits of contact stabilization are achieved if the organic load is present mainly in a colloidal state. Generally, the greater the fraction of soluble BOD, the greater the required contact time. As a result, the required aeration volume of this process approaches that of the conventional process as the relative amount of soluble BOD in the wastewater increases.

Ten-States Standards, however, specify significantly higher contact and stabilization times than those previously cited, especially for the smaller sized plants as indicated in Table 5-5 (10).

Table 5-5

Suggested Design Guidelines

Plant Design Flow mgd	Contact Time[1] hours	Stabilization Time[2] hours
to 0.5	3.0	6.0
0.5 to 1.5	3.0 to 2.0	6.0 to 4.0
1.6 and up	2.0 to 1.5	4.0 to 3.0

[1] Based on average design forward flow
[2] Based on average design recycle flow

These values were no doubt selected to compensate for the extreme flow variations that occur at small treatment plants. However, when Ten-States Standards design criteria are applied to smaller plants, they may result in poor quality effluents (11). McKinney (12) has indicated that, in typically designed contact stabilization plants, all of the stabilization of the organic matter in the raw wastewater occurs in the contact zone; therefore, only endogeneous respiration occurs in the stabilization tank. This situation results in partial stabilization of the sludge in the contact tank, which causes poor settling characteristics in the secondary clarifier.

5.2.4 Completely-Mixed Activated Sludge

In the past at many small activated sludge package plants, oxygen was supplied by mechanical aerators, which provided nearly completely-mixed conditions. However, the specific advantages of the completely-mixed system are just recently being realized (13) (14).

One of the main advantages of the completely-mixed process is related to the introduction of influent waste and the recycled sludge uniformly throughout the aeration tank, as indicated in Figure 5-5. This allows for uniform oxygen demand throughout the aeration tank. This flow pattern also adds some operational stability when treating slug loads of industrial wastes. Operational data from four plants utilizing complete-mix are presented in Table 5-6.

FIGURE 5-5

COMPLETELY-MIXED FLOW DIAGRAM

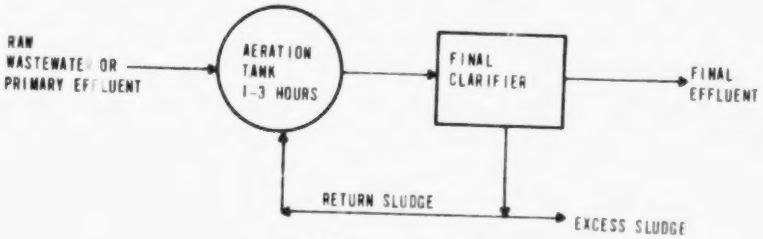


Table 5-6

Operational Data from Various Completely-Mixed Activated Sludge Plants

Plant Location	Influent Flow	Sludge Recycle	BOD		Aeration Tank MLSS	Organic Loading	Volumetric Loading	Aeration Detention Time[1]	Air Supplied per lb. of BOD Removed	Secondary BOD Removal Efficiency	Reference
			Influent	Effluent							
	mgd	percent	mg/l	mg/l	mg/l	lbs. BOD/day / lb. MLSS	lbs. BOD/day / 1,000 cu ft	hours	cu ft./lb.	percent	
Nebraska	3.4	50	250	15	4,500	0.27	80	5.0	500	94	13
Nebraska	4.1	100	270	13.5	4,500	0.32	97	4.4	500	95	13
Nebraska	5.0	200	280	6	4,500	0.38	116	3.8	560	98	13
Nebraska	0.38	26	225	25	4,230	0.48	126	2.6	–	89	15
Nebraska	0.43	40	227	32	5,460	0.42	142	2.5	–	86	15
Texas	0.29	82	115	9	3,820	0.21	50	3.7	–	92.5	16
Texas	0.29	100	141	25	5,000	0.20	62	3.7	–	82.0	16
Texas	0.30	145	123	19	5,540	0.16	54	2.2	–	83.0	16
Texas	0.37	100	180	17	5,620	0.29	103	3.0	–	91.0	16
Illinois	1.6	21	102	8	6,500	0.17	74	2.20	1,670	92.5	17
Illinois	1.94	21	80	13	6,000	0.195	73	1.75	1,900	84.0	17
Illinois	1.91	25	80	19	6,500	0.18	72	1.80	1,380	76.0	17
Illinois	1.55	25	108	18	6,300	0.20	79	2.2	1,290	83.0	17

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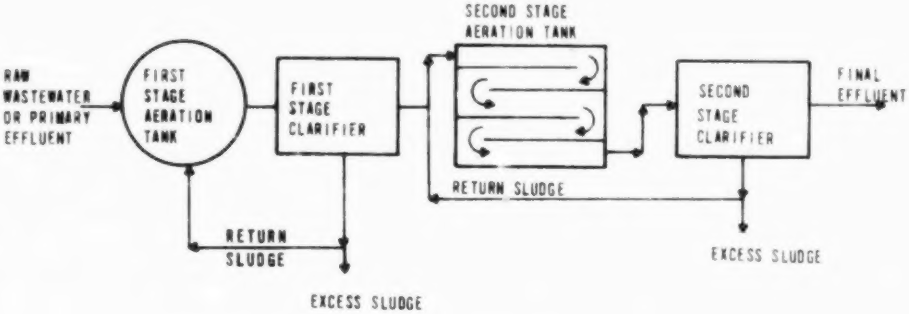
[1] Excluding sludge recycle.

5.2.5 Two-Stage Activated Sludge

Two-stage activated sludge is essentially two separate activated sludge processes operating in series, as shown in Figure 5-6. One of the chief advantages of this flow scheme is in the area of nitrification. The two separate sludge systems permit the development of two specialized microbial populations. In the first stage, the bulk of the carbonaceous material is removed by a wide variety of heterotrophic organisms commonly found in activated sludge. The reduction of BOD in the first stage permits an accumulation of the slower growing nitrifying oragnisms in the second stage which oxidize the ammonia nitrogen to the nitrate form.

FIGURE 5-6

TWO-STAGE ACTIVATED SLUDGE FLOW DIAGRAM


SECOND STAGE AERATION TANK
RAW WASTEWATER OR PRIMARY EFFLUENT
FIRST STAGE AERATION TANK
FIRST STAGE CLARIFIER
SECOND STAGE CLARIFIER
FINAL EFFLUENT
RETURN SLUDGE
RETURN SLUDGE
EXCESS SLUDGE
EXCESS SLUDGE

Although attention has recently been given to this modification, it is generally not considered economical for upgrading unless nitrification is a major consideration. This point is emphasized by examining the operational data presented in Table 5-7. The incremental BOD removal obtained in the second stage generally is not competitive with alternative carbonaceous removal options unless nitrification is also required. The advantage of satisfying the oxygen demand of the ammonia nitrogen normally discharged should not be underestimated since this can amount to as high as 70 percent of the total oxygen demand of the plant secondary effluent (20). Note the high air requirement in lbs. air supplied/lb. BOD removed in the second stage (Table 5-7). A portion of this air is used for ammonia oxidation.

5.2.6 Oxygen Aeration

The Linde Division of the Union Carbide Corporation recently introduced an activated sludge system utilizing oxygen instead of air and termed it the UNOX process. Subsequently, several other companies have introduced oxygen aeration contacting systems.

Table 5-7

Operational Data of Two-Stage Activated Sludge Plants

Plant Location	Influent Flow mgd	1st Stage Performance						1st Stage BOD Removal Efficiency percent	Air Supplied per lb. of BOD Removed cu ft /lb	Aeration Detention Time[1] hrs
		Sludge Recycle percent	BOD Influent mg/l	BOD Effluent mg/l	Aeration Tank MLSS mg/l	Organic Loading lbs. BOD/day / lb. MLSS	Volumetric Loading lbs. BOD/day / 1,000 cu ft			
Pennsylvania	2.27	27	204	35.5	3,760	0.86	182	82.2	1,180	1.6
Pennsylvania	2.24	19	220	22.7	2,020	1.56	197	89.2	1,180	1.7
Pennsylvania	1.47	48	271	32.4	2,350	1.10	160	87.3	1,180	2.5
Pennsylvania	1.28	46	249	32.8	1,980	1.02	128	83.4	1,180	2.9
Pennsylvania	2.10	27	223	40.5	1,920	1.56	188	81.8	1,180	1.8
Pennsylvania	0.21	56	138	12.0	3,150	1.0	205	95.8	820	0.7
Pennsylvania	0.21	56	266	29.0	3,150	1.9	375	89.0	820	0.7
Pennsylvania	0.21	56	104	19.0	2,650	0.92	154	81.8	820	0.7
Pennsylvania	0.21	56	133	25.0	2,650	1.20	196	81.4	820	0.7
Pennsylvania	0.21	56	110	13.0	2,650	0.95	158	88.5	820	0.7
Pennsylvania	0.21	56	134	10.0	2,650	1.10	186	92.7	820	0.7

[1] Excluding sludge recycle.

Table 5-7
(continued)

2nd Stage Performance											
Sludge Recycle	BOD		Aeration Tank MLSS	Organic Loading	Volumetric Loading	2nd Stage BOD Removal Efficiency	Air Supplied per lb. of BOD Removed[2]	Aeration Detention Time[3]	Overall Secondary BOD Removal	Reference
	Influent	Effluent								
percent	mg/l	mg/l	mg/l	lbs. BOD/day / lb. MLSS	lbs. BOD/day / 1,000 cu.ft.	percent	cu.ft./lb	hours	percent	
24	36	12	1,520	0.34	33.0	67.0	4,600	1.6	94.0	18
11	23	11	1,600	0.21	21.0	53.7	4,600	1.7	95.0	18
23	32	19	820	0.36	19.0	40.4	4,600	2.5	93.0	18
29	33	19	800	0.34	17.0	42.5	4,600	2.9	93.0	18
14	41	17	935	0.59	35.0	59.0	4,600	1.8	93.0	18
19	12	7	1,500	0.21	20.5	41.6	4,600	---	95.0	19
19	29	8	1,510	0.47	44.0	72.2	4,100	0.7	96.0	19
19	19	4	1,350	0.35	29.0	79.0	4,100	0.7	96.0	19
19	25	21	1,350	0.49	41.5	16.0	4,100	0.7	81.0	19
19	13	8	1,350	0.24	20.5	38.5	4,100	0.7	93.0	19
19	10	7	1,350	0.20	17.0	30.0	4,100	0.7	95.0	19

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[2] Including that needed for ammonia oxidation.

[3] Excluding sludge recycle.

A schematic diagram of the UNOX system is shown in Figure 5-7 (21). In the oxygen aeration process, the aeration tank is staged by using baffles, and is completely covered to provide a gas-tight enclosure. The influent wastewater, recycled sludge, and oxygen gas are introduced into the first stage, and then flow to subsequent stages. The oxygen is produced at the plant site by either a cryogenic unit or, in the case of smaller plants, a molecular sieve device. A liquid oxygen storage unit is generally recommended to eliminate the duplication of units usually specified by most State Health Departments.

Table 5-8 summarizes operational data from two plants utilizing oxygen aeration. The following is a list of possible advantages of the oxygen aeration process (22):

1. Reduced capital cost.
2. Reduced operating cost.
3. Reduced sludge production.
4. More reliable process control.
5. More effective odor control.
6. Reduced land area.
7. High D.O. in the final effluent.

Of course, the potential economic advantages are a function of local factors and should be confirmed in comparison with other upgrading alternatives.

5.3 Activated Sludge Design Considerations

Initially, plant operators, through a trial and error procedure, developed the most efficient operating criteria for conventional activated sludge plants as well as for the modifications of the process. Out of this evolution, basic design criteria were developed. These criteria are still in use today and, in many cases, are rigidly adhered to by regulatory agencies.

A limitation of these design criteria is that volumetric loading (lbs. BOD/1,000 cu.ft./day) has been considered preferable, for design purposes, to organic loading considerations (lbs. BOD/day/lb. MLVSS). Many of the modifications have shown that organic loading is an important consideration; in fact, a higher volumetric loading has been achieved for modifications of the conventional process at the same organic loading used in the conventional process.

Basic parameters of interest in the design of an activated sludge process are:

1. BOD removal for specific operating conditions.
2. Oxygen (air) requirements for synthesis of organisms and for endogenous reactions.
3. Sludge production.
4. Oxygen transfer rates in wastewater.
5. Nutrient requirements.
6. Separation and return of activated sludge.

FIGURE 5-7

SCHEMATIC DIAGRAM OF MULTI-STAGE OXYGEN AERATION SYSTEM (21)

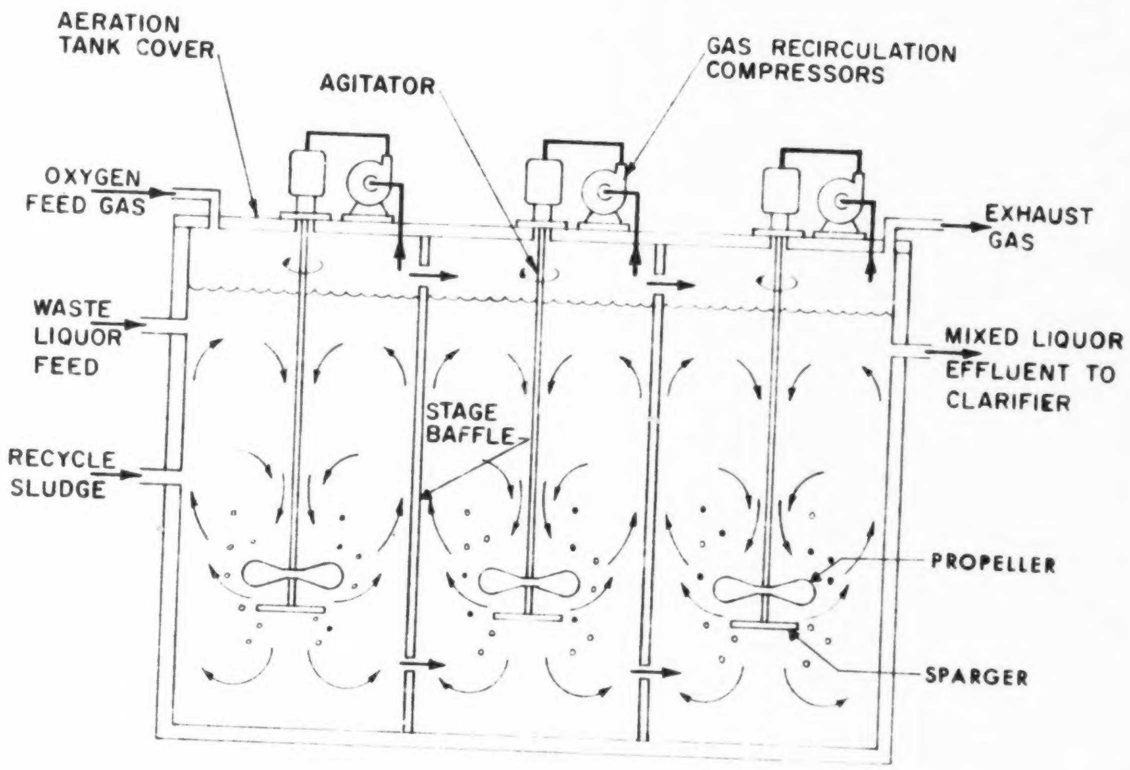

AERATION TANK COVER
AGITATOR
GAS RECIRCULATION COMPRESSORS
OXYGEN FEED GAS
EXHAUST GAS
WASTE LIQUOR FEED
MIXED LIQUOR EFFLUENT TO CLARIFIER
STAGE BAFFLE
RECYCLE SLUDGE
PROPELLER
SPARGER

Table 5-8

Operational Data from Two Oxygen-Aeration Activated Sludge Plants

Plant Location	Influent Flow mgd	Sludge Recycle percent	BOD Influent mg/l	BOD Effluent mg/l	Aeration Tank MLSS mg/l	Organic Loading lbs BOD/day / lb MLSS	Volumetric Loading lbs BOD/day / 1,000 cu ft	Aeration Detention Time[1] hours	Secondary BOD Removal Efficiency percent	Reference
New York	1.33	53	237	22	5,890	0.35	126	2.9	91.0	21
New York	1.29	54	221	18	6,810	0.27	115	3.0	92.0	21
New York	1.38	56	249	19	6,840	0.38	140	2.8	93.0	21
New York	1.19	45	283	19	5,890	0.37	132	3.3	91.5	21
New York	1.36	42	270	9	7,400	0.31	142	2.9	97.0	21
New York	1.41	38	304	11	5,700	0.47	166	2.8	97.0	21
New York	1.64	32	269	15	5,560	0.50	170	2.4	94.5	21
Washington, D.C.	0.07	50	115	19	4,140	0.31	80	2.2	84.0	5
Washington, D.C.	0.10	31.5	102	12	6,000	0.25	90	1.7	88.2	5
Washington, D.C.	0.10	38	116	14	8,120	0.22	108	1.7	88.0	5

---

[1] Excluding sludge recycle.

5.3.1 BOD Removal Rates for Specific Operating Conditions

Eckenfelder (23) has indicated that a linear arithmetic relationship exists between BOD removal rate (mg BOD/hr./gm VSS) and effluent BOD (mg/l), as shown in Figure 5-8 for typical data from various completely-mixed activated sludge plants. The variations in the BOD removal relationship in Figure 5-8 are influenced by the presence of various proportions of domestic and industrial wastes.

Using Figure 5-8, the detention time required to achieve a specific effluent BOD can be obtained as follows:

$$t = \frac{L_a - L_e}{S_a r'}$$

where:

t = Detention time, hours
L_a = Influent BOD to aeration tank, mg/l
L_e = Clarifier effluent BOD, mg/l
S_a = Mixed liquor volatile suspended solids (MLVSS), mg/l
r' = BOD removal rate, mg BOD/hr./gm VSS

Weston (24) developed a log-log relationship between a BOD removal rate constant (r) and a loading ratio (L_o/S_o), defined by the equation:

$$r = \frac{L_o - L_e}{t_e L_e}$$

where:

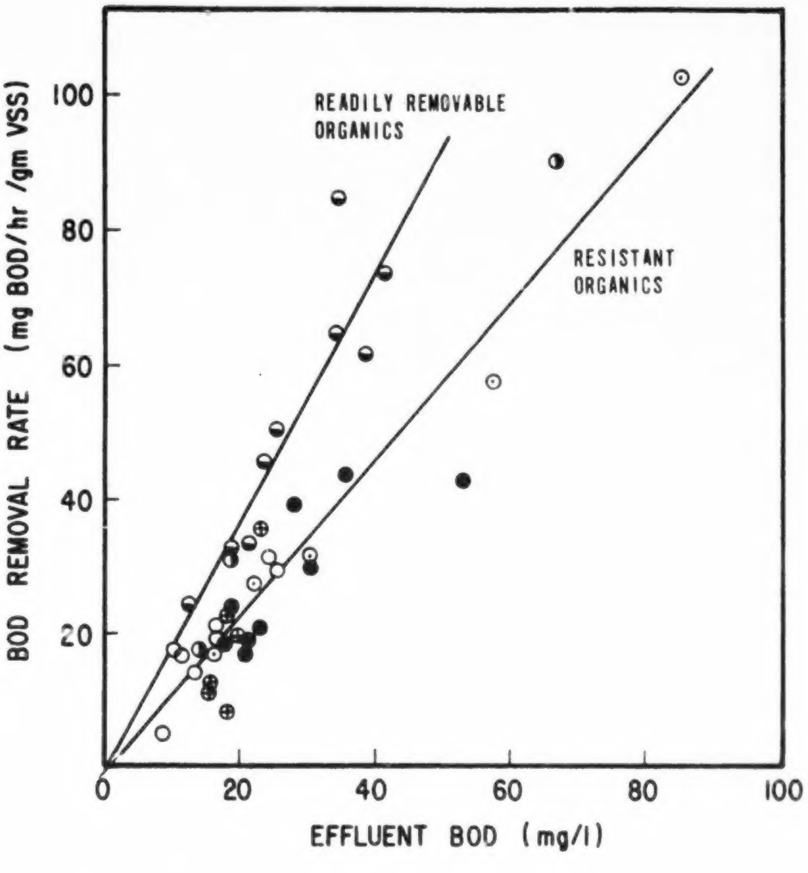
$$L_o = \frac{L_i + RL_e}{1 + R}$$

r = BOD removal rate constant, min^{-1}
L_o = BOD of wastewater after mixture of raw wastewater or primary effluent with sludge recycle, mg/l
L_e = Clarifier effluent BOD, mg/l
t_e = Aeration tank detention time, minutes (including recycle)
L_i = Raw wastewater or primary effluent BOD, mg/l
R = Sludge recycle as percent of influent flow
S_o = MLVSS, mg/l

FIGURE 5-8

BOD REMCVAL CHARACTERISTICS

FOR VARIOUS COMPLETELY MIXED ACTIVATED SLUDGE PLANTS (23)



The aeration tank detention time is related to process efficiency (E) and BOD removal rate (r) by the following:

$$t_e = \frac{E}{100-E} \times \frac{1}{r}$$

where:

$$E = \frac{L_o - L_e}{L_o} \times 100$$

Operational data from over 20 plants with various activated sludge modifications were analyzed using the Weston procedure to determine BOD removal rate constants. Operational data for this analysis were taken from references (2) through (8), (11), (13) through (19), (21), and (25) through (27). The results are summarized in Figure 5-9. It should be pointed out that the BOD removal rate curves represent only average kinetics with no temperature correction applied, and the various loading ratios were determined using MLSS, not MLVSS. For these reasons, these curves are not recommended for design purposes, but are included merely to illustrate the relative kinetic rates of the modifications. The presence of significant quantities of industrial wastes, which may have different removal rate characteristics, would modify or displace the curves shown.

Note that each modification in Figure 5-9 except step aeration results in an increased r and, therefore, a subsequent decrease in detention time (assuming the efficiency remains constant). The basic reason for the non-consistency in the step aeration data is that the S_o value (MLSS) is the average value in the aeration tank. In the step aeration process, the concentration decreases markedly as it proceeds through the aeration tank, but the average concentration is quite similar to the conventional process. Therefore, for the design of step aeration systems, volumetric loading can probably be used, as will be discussed subsequently.

The BOD removal rate curve for the partially treated wastewater in Figure 5-9 represents data obtained from the second stage of three different two-stage biological treatment plants (18) (19) (27). Two of these plants use activated sludge as the first stage, while the third plant uses trickling filtration. The BOD removal rates for the partially treated wastewaters are markedly lower than those of conventional activated sludge because the organics remaining in the first-stage effluent are more resistant to biological degradation than those entering a conventional plant.

Figure 5-10 was prepared using the same operational data as used for Figure 5-9 to show a correlation between volumetric and organic loading rates. The results shown in Figure 5-10 indicate that, for the same organic loading, the volumetric loading increases for various modifications of the conventional process, thereby reducing the required aeration volume.

FIGURE 5-9

RELATIONSHIP BETWEEN BOD REMOVAL RATE CONSTANTS
AND LOADING RATIOS
FOR VARIOUS ACTIVATED SLUDGE MODIFICATIONS

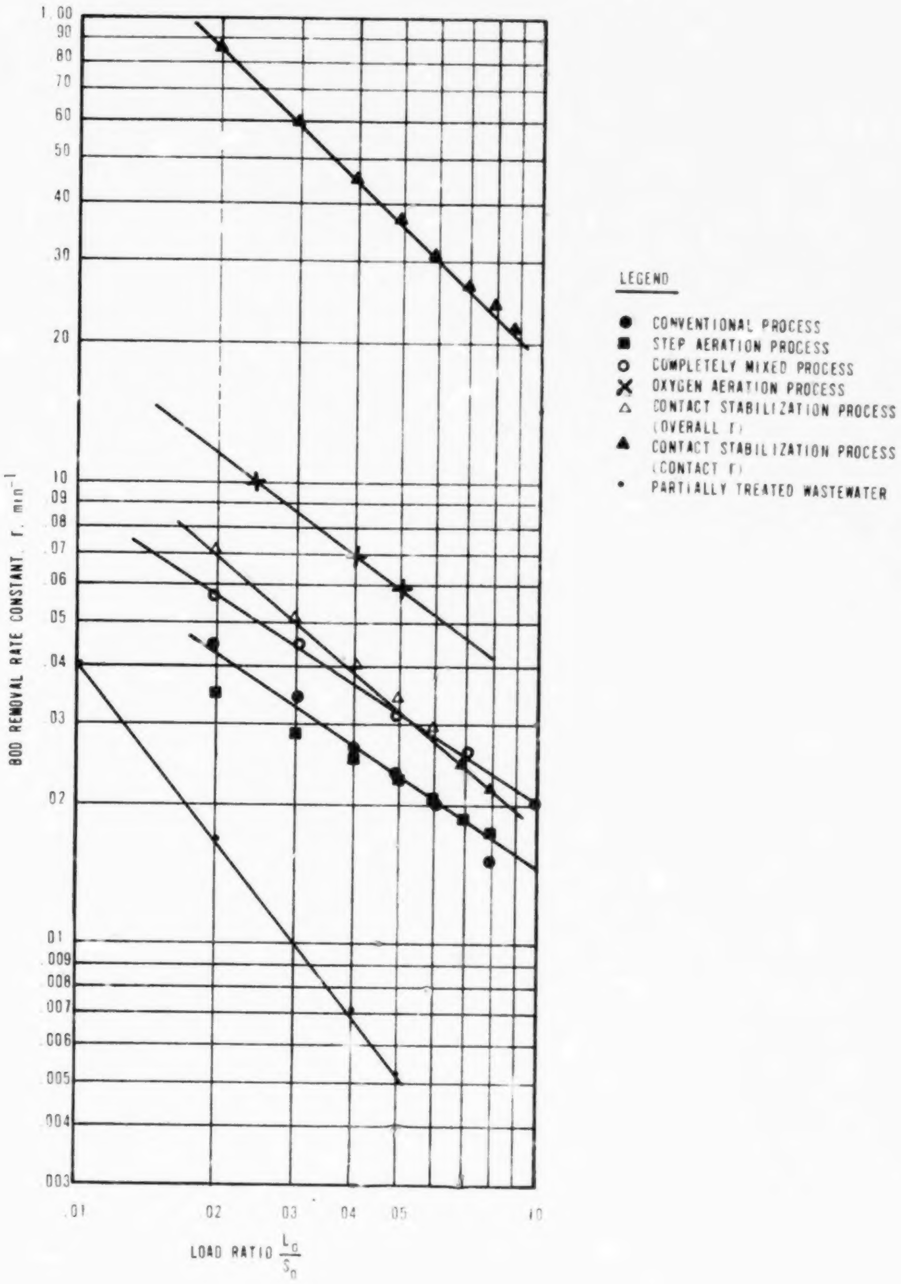
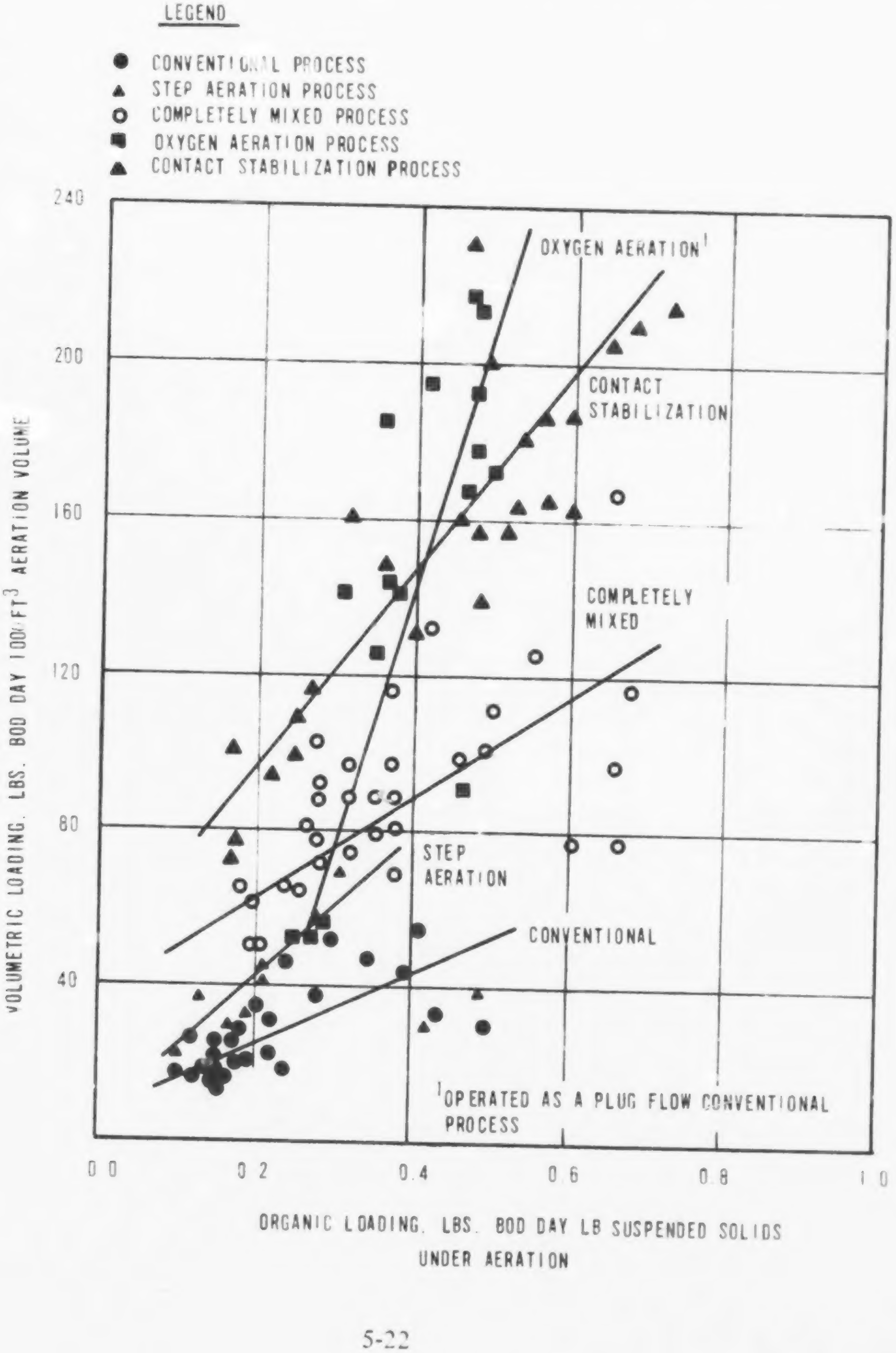


FIGURE 5-10

RELATIONSHIP BETWEEN VOLUMETRIC AND ORGANIC LOADINGS FOR VARIOUS ACTIVATED SLUDGE MODIFICATIONS



Various relationships between BOD removal and organic loading were developed using the same operational data that were used to prepare Figures 5-9 and 5-10, and these relationships are presented in Figure 5-11. Contact stabilization, completely mixed, and the oxygen aeration modifications generally show a slightly higher percent BOD removal at the same loading than does the conventional activated sludge process.

5.3.2 Air Requirements for Synthesis of Organisms and for Endogenous Reactions

Table 5-9 contains ranges for the cubic feet of air required per pound of BOD removed for the various activated sludge modifications previously discussed.

Table 5-9

Air Requirements for
Various Activated Sludge Modifications

Process	Standard cu.ft. Air/lb. BOD Removed
Conventional	1,000 to 700
Step Aeration	700 to 500
Contact Stabilization	750
Completely-Mixed	600

Source: Eckenfelder (23)

5.3.3 Sludge Production

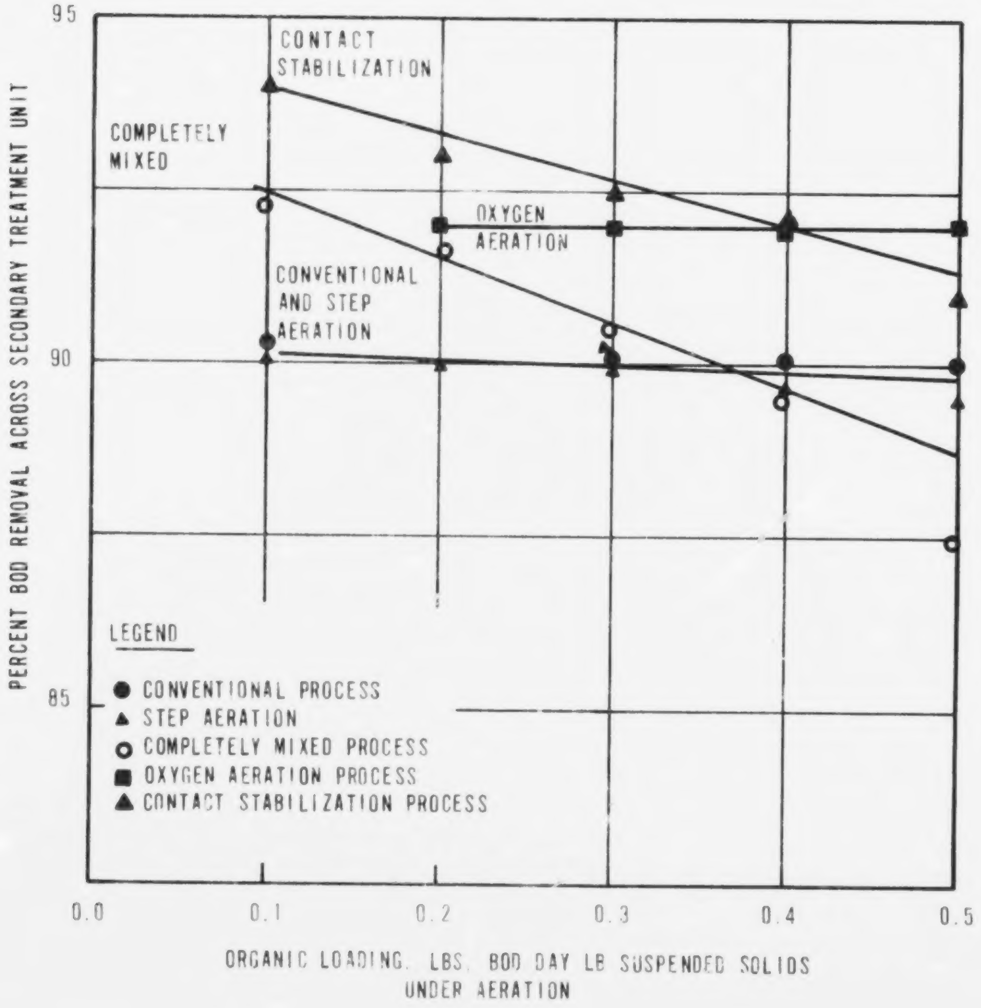
Normally, the activated sludge processes generate excess sludge in relation to the organic loading maintained in the system (9). For common ranges of organic loadings, namely 0.3 to 0.6 lb. BOD/lb.MLVSS/day, it has been observed that the amount of excess sludge produced in the conventional and various modified processes generally varies between 0.5 and 0.7 lb. VSS/lb. BOD removed (6) (13) (28). in contrast to these observations, preliminary results obtained using oxygen aeration indicate that, for organic loadings of 0.4 to 0.8 lb. BOD/lb. MLVSS/day, the excess sludge production was 0.3 to 0.45 lb. VSS/lb. BOD removed (28).

5.3.4 Oxygen Transfer Rates in Wastewater

Oxygen transfer rates in wastewater are affected by various physical and chemical variables, e.g., temperature, degree of turbulent mixing, liquid depth in the aeration tank, oxygen composition of aerating gas, type of aeration device, and chemical characteristics of the wastewater. The major area often overlooked in the past by design engineers has been the effect of an industrial waste on the overall oxygen transfer rate of a system. Where

FIGURE 5-11

RELATIONSHIP BETWEEN BOD REMOVAL AND ORGANIC LOADING FOR VARIOUS ACTIVATED SLUDGE MODIFICATIONS



the industrial waste makes up a large proportion of the total flow it is desirable to verify oxygen transfer rates in the laboratory for proper sizing of aeration units. Oxygen transfer capability of several aeration devices are indicated in Table 5-10.

Table 5-10

Comparison of the Aeration Costs of Various Systems

Type of Aeration System	lb. O_2/hp/hr[1]	kwh/lb. O_2	Power[2] Cost/Day	Relative Power Cost	Relative Capital Cost
Diffused-Air, Fine-Bubble	2.1	0.35	1.26	1.75	3.2
Diffused-Air, Coarse-Bubble	1.4	0.55	1.98	2.75	2.5
Mechanical Aeration, Vertical Shaft	3.7	0.20	0.72	1.0	1.0
Agitator Sparged System	2.1	0.35	1.26	1.75	2.25

[1] Oxygen transfer capabilities shown are for standard conditions in tap water, i.e. 20°C, 760 mm barometric pressure, and initial dissolved oxygen equal to 0 mg/l.

[2] Dollars/lb. O_2/day based on 1.5 cents/kwh power cost.

Source: Mechanical Aeration Seminar (31)

5.3.5 Nutrient Requirements

It is necessary that sufficient nitrogen and phosphorus be present in a wastewater such that neither nutrient becomes the limiting factor in microbial growth reactions encountered in the activated sludge process. Normally, supplemental nutrients are not required for municipal wastewater treatment plants because adequate quantities are available in domestic wastewaters to make organic carbon the limiting macronutrient. For optimum operation of the activated sludge process, the minimum ratio of raw wastewater BOD:N:P is 60:3:1 (29).

5.3.6 Separation and Return of Activated Sludge

Basically, the ability of activated sludge to be separated in a final clarifier does not change appreciably for the various modifications. Clarifier requirements are, therefore, essentially the same, regardless of the modification implemented, provided that operating conditions remain the same. It is for this reason that the Ten-States Standards recommends an average surface overflow rate of 800 gpd/sq.ft. for all of the activated sludge modifications previously discussed, except contact stabilization, when the design capacity is over 1.5 mgd. For the contact stabilization process, the Ten-States Standards recommends an average surface overflow rate of 700 gpd/sq.ft., because with this modification primary sedimentation is often omitted.

In the past, the importance of the final clarifier as an integral unit of the activated sludge process has not been fully recognized. However, recent work by Dick (30) indicates that the final clarifier design is an important aspect in the design of an activated sludge process and that improper clarifier design is often the cause of inefficient BOD and suspended solids removal. He recommends that both aspects of clarification and thickening be considered in the design of final clarifiers.

The recent appreciation of the importance of the final clarifier on activated sludge process efficiency has raised doubt as to the advisability of designing clarifier overflow rates solely on the basis of average or nominal design flow. Many engineers now prefer to size clarifiers on the basis of maximum daily flow. This technique provides greater protection against system solids washout, at the expense of a somewhat larger clarifier. Depending on wastewater characteristics, geometric configuration, and pretreatment considerations, a range of 1,000 to 1,600 gpd/sq.ft. is suggested as a guideline for the maximum allowable surface overflow rate in an activated sludge final clarifier.

Control of sludge recycle is probably the most important operational tool the plant operator has at his disposal to intellegently manage the sludge inventory. Therefore, it is extremely important to provide sufficient sludge recycle capacity to give the operator the required operating flexibility to handle the highly variable and fluctuating waste loads characteristic of many plants.

Two techniques which have been used to control sludge recycle are:

1. Automatically varying the recycle flow to maintain a set relationship to influent flow.

2. Controlling the recycle pumps by a sludge blanket sensor to maintain a predetermined blanket height in the final clarifier.

A firm sludge recycle capacity of at least 50 percent is recommended for the conventional and step aeration processes; at least 100 percent is recommended for the contact stabilization and completely-mixed modifications. Firm capacity is defined as the available pumping capacity with the largest pump out of service.

5.4 Pilot Studies

The use of pilot facilities for investigating the upgrading of existing activated sludge plants is strongly indicated in many cases, to ensure that optimum design parameters are selected. There are two general types of piloting facilities available: batch or continuous-flow systems. Continuous-flow systems may range in size from bench-scale to 5 or 10-gpm units. The basic objective of either a batch or a continuous-flow study is to generate parameters necessary for design. Some of the parameters of basic interest are BOD removal rates, oxygen requirements, and sludge production.

5.4.1 Batch Studies

Batch laboratory-scale units are subject to all of the inherent difficulties of biological oxidation systems, with the added magnified complexities of large surface-to-volume ratios, small quantities of sludge mass in the reactor, and the undesirable factors associated with slug feeding of wastewater. In spite of these inherent difficulties, batch studies have attractive features in that they afford an economic and efficient controlled method of developing fundamental information concerning the applicability of various activated sludge modifications. However, use of the continuous-flow system is preferable to obtain design parameters since it approximates the operation of an actual plant, permitting evaluation of the effects of variations in treatability characteristics, as well as of variations in wastewater loading or strength.

5.4.2 Continuous-Flow Studies

Continuous-flow units, as opposed to batch units, must be used on wastewaters which exhibit biostatic or exert toxic effects. However, most of the municipal waters do not exhibit this property unless there is a significant discharge of untreated industrial wastes.

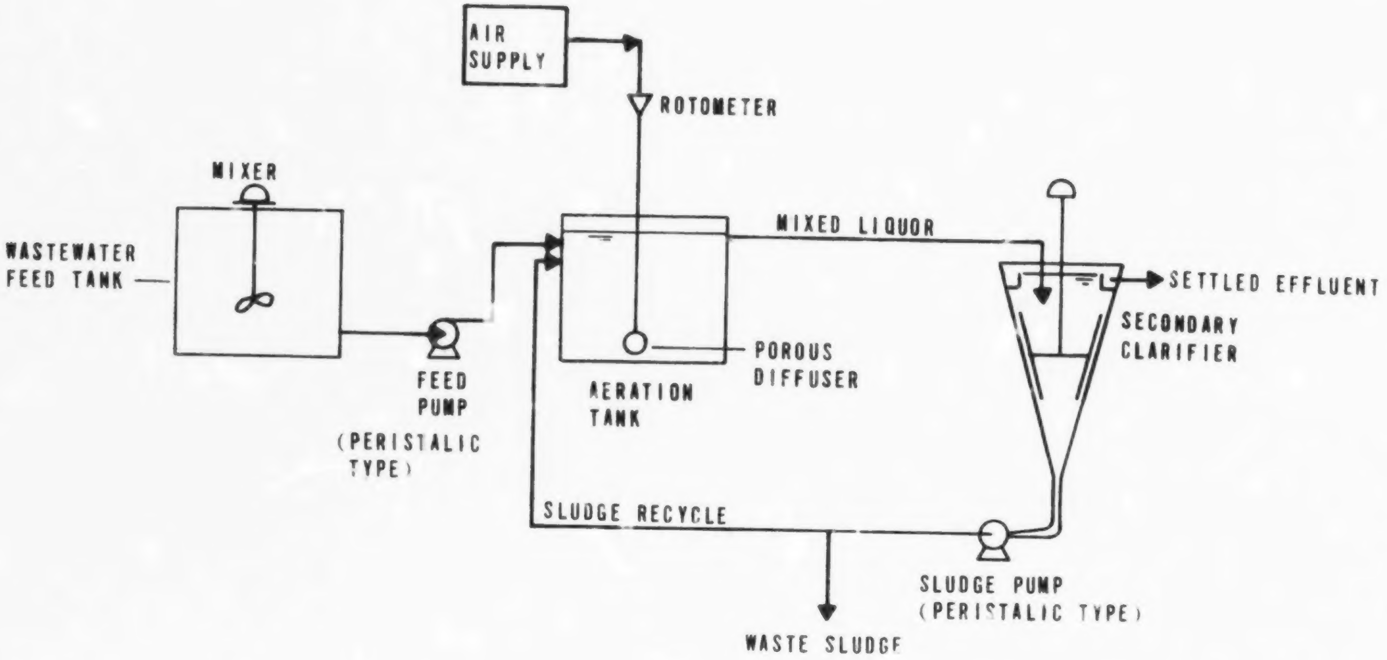
A schematic diagram of a continuous-flow aeration unit is shown in Figure 5-12. Basically, the system consists of a wastewater feed tank provided with a mixer to blend the wastewater, prior to feeding to the aeration unit. The use of a mixer in addition to blending prevents solids deposition in the feed tank. The wastewater from the feed tank and the recycle sludge from the secondary clarifier are pumped to the aeration tank using peristalic type pumps. The aeration in laboratory units is normally supplied through porous diffusers, and the air supply is controlled by the use of rotameters. The wastewater, after treatment, flows by gravity to a clarifier where the mixed liquor solids are separated. In the laboratory-scale clarifier, care must be taken to prevent solids deposition on the side walls of the clarifier. To accomplish this, the clarifier should have a scraper mechanism which aids in both the settling and removal of the mixed liquor solids. If 24-hr. composite sampling of feed wastewater and clarifier effluent are required, provision should be made to pump these streams into refrigerated sample bottles.

The selected size of the aeration system depends on the wastewater strength and the desired detention time. Slow pumping rates are difficult; therefore, for longer detention times, larger aeration volumes are desirable. Also, higher organic loadings due to high strength wastes require a larger aeration volume.

Two approaches may be applied for the acclimation and growth of a culture of microorganisms for use in a continuous-flow system. An available activated sludge culture may be utilized as the source of microorganisms, with the normal feed to that system being gradually replaced by the wastewater under investigation until satisfactory performance on that wastewater is obtained. Alternatively, culture development can begin with a small quantity of seed organisms and a wastewater feed diluted below the toxicity

FIGURE 5-12

SCHEMATIC OF A CONTINUOUS-FLOW AERATION UNIT


AIR SUPPLY
ROTOMETER
MIXER
WASTEWATER FEED TANK
MIXED LIQUOR
SETTLED EFFLUENT
SECONDARY CLARIFIER
POROUS DIFFUSER
FEED PUMP
AERATION TANK
(PERISTALIC TYPE)
SLUDGE RECYCLE
SLUDGE PUMP
(PERISTALIC TYPE)
WASTE SLUDGE

threshold (if toxicity exists). As the biological mass develops, the toxicity threshold is redetermined and wastewater concentration is increased accordingly until the culture is capable of handling wastewater at 100 percent concentration. The latter technique is preferred because it provides the best opportunity to observe the growth characteristics of the biological culture as well as potential problems with acute or chronic toxicity.

When the culture is capable of functioning on the undiluted wastewater, data are collected on the performance of the system, beginning with a low-feed rate and increasing the feed rate until performance near that for the anticipated design is achieved. For various organic loadings (lbs. BOD/day/lb. MLVSS), the performance and characteristics of the system should be evaluated in terms of:

1. BOD removal.
2. COD removal.
3. Oxygen consumption.
4. Concentration of biological solids.
5. Characteristics of biological culture (microscopic appearance and settling rates).
6. Physical nature of the effluent (suspended solids, odor, color, etc.).

5.4.2.1 BOD Removal Rate Determinations

The data collected from continuous-flow units can be analyzed using either the Eckenfelder or Weston procedures in order to define the appropriate BOD removal rate constant for design conditions as previously discussed (23) (24).

5.4.2.2 Oxygen Uptake Requirements

The oxygen consumption data obtained in the continuous-flow pilot unit are evaluated to obtain energy and endogenous oxygen requirements as shown in Figure 5-13, which is a schematic of an oxygen uptake curve for a typical continuous-flow activated sludge unit (29). The slope of the line (m) represents the oxygen required for cell synthesis, while the ordinate intercept (b) represents the oxygen required for endogenous respiration. The net oxygen consumption is expressed by the following formula:

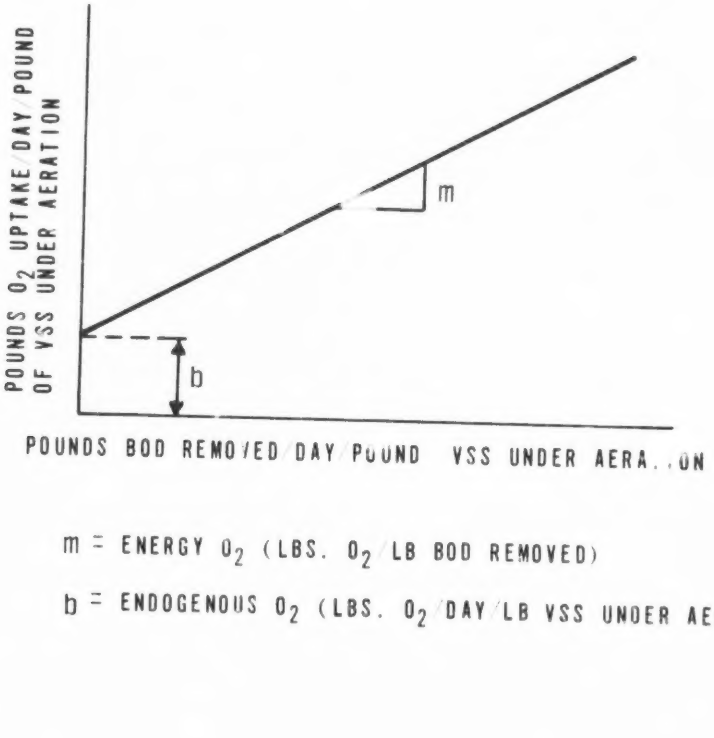
$$O_2 = m \left(\frac{\text{lbs. BOD removed}}{\text{day}} \right) + b \text{ (lbs. VSS under aeration)}$$

where:

O_2 = lbs. oxygen uptake/day
m = energy oxygen, lbs. oxygen uptake/lb. BOD removed
b = endogenous oxygen, lbs. oxygen uptake/day/lb. VSS under aeration

FIGURE 5-13

DETERMINATION OF OXYGEN UPTAKE REQUIREMENTS (29)



m = ENERGY O_2 (LBS. O_2/LB BOD REMOVED)

b = ENDOGENOUS O_2 (LBS. O_2/DAY/LB VSS UNDER AERATION)

5.4.2.3 Sludge Production

Sludge production in an activated sludge system is expressed as the net effect of two processes as follows:

1. A production of new organisms resulting from the synthesis of a portion of the organic material removed.

2. A reduction of the weight of organisms under aeration by the process of self-oxidation or endogenous respiration.

Figure 5-14 is a schematic representation of sludge production from a continuous-flow pilot plant (29). The slope of the line (m) represents sludge synthesis, while the ordinate intercept (b) represents the endogenous destruction of solids. The net sludge production is expressed by the following equation:

$$\text{VSS produced/day} = m^1 \text{ (lbs. BOD removed/day)} - b^1 \text{ (lbs. VSS under aeration)}$$

where:

m^1 = sludge synthesis (lbs. VSS produced/lb. BOD removed)
b^1 = endogenous destruction of sludge (lbs. VSS destroyed/day/lb. VSS under aeration)

Sludge production, like any other biological process, is temperature dependent. Therefore, sludge production data obtained from a pilot study must be adjusted for the temperature ranges which are to be experienced by the full-scale plant.

5.5 Activated Sludge Upgrading Techniques and Design Basis

Upgrading of conventional activated sludge plants may be required because the plants are hydraulically and organically overloaded, because of the need for increased treatment efficiency, or both. Upgrading to relieve overloaded conditions and upgrading to improve removal efficiency to meet higher water quality standards are covered in the following two sections.

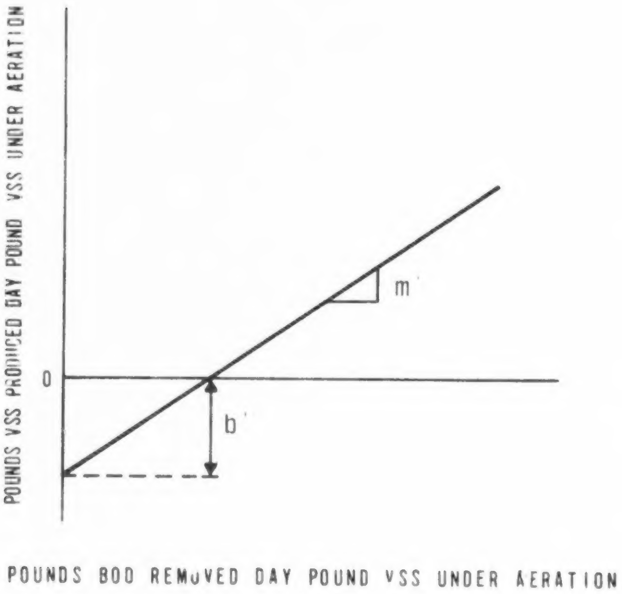
5.5.1 Upgrading to Relieve Organic and Hydraulic Overloading

The following activated sludge modifications are examined as they apply to the upgrading of an existing overloaded conventional activated sludge plant:

1. Step aeration and contact stabilization - these processes are combined because of their similarities.
2. Completely-mixed.
3. Oxygen aeration.
4. Use of activated sludge to treat partially treated wastewater.

FIGURE 5-14

DETERMINATION OF SLUDGE PRODUCTION CHARACTERISTICS (29)



m' = SLUDGE SYNTHESIS (LBS. VSS LB BOD REMOVED)

b' = ENDOGENOUS DESTRUCTION OF SLUDGE (LBS. VSS DAY POUND VSS UNDER AERATION)

Before examining each individual upgrading procedure, several general statements can be made. Operational data, BOD removal rate constants, and volumetric loadings previously discussed indicate that all of the activated sludge modifications are applicable for upgrading an overloaded conventional activated sludge plant. These modifications will require consideration of renovating the air system to supply more air per unit of aeration tank volume. This is because the volumetric loadings for various modifications are substantially higher than for the conventional process as shown in Figure 5-10, even though the air requirements per pound of BOD removed decrease slightly as indicated in Table 5-9.

Therefore, to upgrade using one of the previously discussed modifications, it will generally be necessary to install an air system which will be capable of supplying air at a higher rate than was previously required by the conventional process. This may be done either by enlarging the existing air supply facilities, or by adding surface or mechanical aerators where applicable.

Table 5-10 contains a comparison of capabilities and costs of various aeration systems commonly employed in the activated sludge process (31). The mechanical aerator and agitator sparger systems are illustrated in Figure 8-2 in the Post-Aeration Section. The data presented in Table 5-10 indicate the increased oxygen transfer capability and the lower capital and operating costs for the mechanical aerator. Even though mechanical aerators afford a high transfer efficiency, their use in an existing basin may pose problems because the geometric configurations required for their most efficient use may be quite different than the existing basin configuration. Most existing conventional plants use either a fine or coarse bubble diffused air system. The fine bubble system is more efficient and cheaper to operate, but on the other hand represents a greater capital investment and a costlier maintenance problem than a coarse bubble system.

The use of mechanical aerators and agitator sparger systems has gained popularity in the recent decade. The aerator has a high oxygen transfer efficiency, but this advantage is partially lost in upgrading a plug flow type aeration tank due to adverse geometric configuration requiring multiple units. This was found to be true for an upgrading investigation performed for the City of Baltimore, Maryland (32). In an economic comparison, it was found that the annual costs for a diffused versus a mechanical aeration system were approximately equal because of the existing configuration of the plug flow basins. However, it was recommended that mechanical aeration definitely be considered for future aeration tank expansion.

The agitator sparger system has an operational advantage over the diffused air unit (coarse or fine bubble) in that during low flows the air may be reduced but the mixing will be maintained due to the action of the turbine agitator.

5.5.1.1 Step Aeration and Contact Stabilization (Examples A and B)

Step aeration has been used successfully as an upgrading technique in New York City; Indianapolis, Indiana; and numerous other locations. Contact stabilization has been used in Austin, Texas; York, Pennsylvania; and Bergen County, New Jersey.

The step aeration and contact stabilization processes have been grouped together because both modifications can be incorporated into the upgraded design at a minimum capital investment. Added flexibility in the use of the two modifications is accomplished by sizing the influent step aeration piping so that the entire flow may be introduced in the last bay of the aeration tank, thus permitting operation as a contact stabilization process.

This type of upgrading, if applicable, generally requires a minimal capital investment. Again, it must be stressed that if the soluble BOD in the wastewater is expected to increase significantly over the design period of the plant due to an increase of industrial wastewater discharged to the municipal plant, then contact stabilization may not be as efficient as other alternatives. Hence, upgrading under this condition should preferentially consider use of the completely mixed or oxygen aeration modifications. Examples A, B, and C are desk-top analyses to illustrate the design considerations involved in upgrading a conventional activated sludge plant to step aeration, contact stabilization, and completely-mixed flow patterns, respectively.

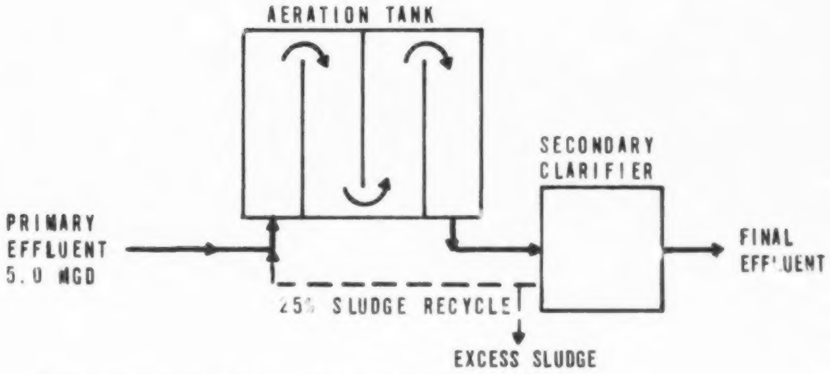
A schematic flow diagram for upgrading a conventional plant to step aeration is presented as Example A in Figure 5-15. A comparison of the original design values for the conventional plant before it was overloaded as well as the data from the overloaded plant before upgrading are shown in Table 5-11. The upgrading of the plant was required due to an increase of flow from 5 to 8.4 mgd, which increased the effluent BOD from 20 mg/l to 35 mg/l.

To upgrade the plant to its previous performance, it was decided to employ step aeration. The design performance of the upgraded plant is also shown in Table 5-11. To implement this upgrading, it was necessary to modify the influent piping, renovate the air system in order to deliver 700 cu.ft. of air/lb. BOD removed, and to expand the average sludge recycle capacity to 2.1 mgd, or 25 percent of the upgraded flow In addition, the secondary clarifier capacity was increased to maintain an average overflow rate of 800 gpd/sq.ft. Due to the increased hydraulic load, the primary clarification capacity and other ancillary operations, such as excess sludge handling and disposal facilities, would also require evaluation for upgrading to match the increased capacity and performance of the activated sludge process. The cost for additional primary clarifier or sludge handling capacity will not be considered in this unit operations section.

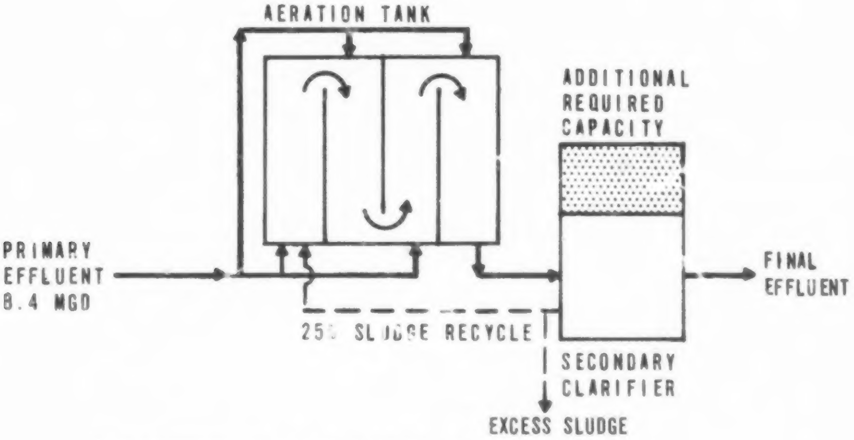
FIGURE 5-15

UPGRADING A CONVENTIONAL ACTIVATED SLUDGE PROCESS TO STEP AERATION

EXAMPLE A


AERATION TANK
SECONDARY CLARIFIER
PRIMARY EFFLUENT 5.0 MGD
FINAL EFFLUENT
25% SLUDGE RECYCLE
EXCESS SLUDGE

TREATMENT SYSTEM BEFORE UPGRADING
CONVENTIONAL ACTIVATED SLUDGE (DIFFUSED AIR SYSTEM)


AERATION TANK
ADDITIONAL REQUIRED CAPACITY
PRIMARY EFFLUENT 8.4 MGD
FINAL EFFLUENT
25% SLUDGE RECYCLE
SECONDARY CLARIFIER
EXCESS SLUDGE

TREATMENT SYSTEM AFTER UPGRADING
STEP AERATION PROCESS

Table 5-11

Upgrading Conventional Activated Sludge to Step Aeration - Example A

Description	Original Design Before Overloading	Overloaded Design Condition	Upgraded Design Condition
Flow, mgd	5.0	8.4	8.4
Influent BOD, mg/l	200	200	200
Primary Treatment			
Percent BOD Removal	30	30[1]	30
Aeration Tank			
MLSS, mg/l	2,000	--	2,000
Sludge Recycle, percent	25	15	25
Air Requirement, cu.ft. air/lb. BOD removed	800	--	700
Volumetric Loading, lbs. BOD/day/1,000 cu.ft.	35	62	62
Organic Loading, lbs. BOD/day/lb. MLSS	0.34	0.88	0.54
Detention Time in Aerator, minutes[2]	300	180	180
Secondary Clarifier			
Overflow Rate, gpd/sq.ft.	800	1,280	800
Secondary Treatment			
Percent BOD Removal	86.0	75.0	86.0
Effluent BOD, mg/l	20	35	20

[1] Requires modification of primary clarifier to handle increased hydraulic load to achieve 30 percent BOD removal.

[2] Excluding sludge recycle.

The capital costs for upgrading these secondary units were estimated at $410,000 ($120 per 1,000 gpd of incremental upgraded capacity) and were allocated as follows:

Aeration Tank Modification	$160,000
Secondary Clarifier Expansion	250,000
TOTAL	$410,000[1]

Example B illustrates upgrading a conventional activated sludge plant equipped with mechanical aerators to contact stabilization. A schematic flow diagram of the plant before and after upgrading is shown in Figure 5-16. Table 5-12 contains design data from the plant while overloaded and after it was upgraded. The plant was upgraded from 1.2 to 3.0 mgd using the contact stabilization modifications. The effluent BOD was upgraded from 40 mg/l to 20 mg/l.

Capital costs for this modification include revamping the influent piping, expanding the sludge recycle to 75 percent of the upgraded flow, and installing new mechanical aerators capable of delivering 3.5 lbs O_2 per hp/hour under standard conditions. In this example, the primary clarifier was incorporated into the secondary clarification facilities, thereby providing an average overflow rate of 780 gpd/sq.ft. To implement these modifications, the capital costs were estimated at $370,000 ($206 per 1,000 gpd of incremental upgraded capacity) and were allocated as follows:

Aeration Tank Modifications	$340,000
Conversion of Present Primary to Secondary Clarifier	30,000
TOTAL	$370,000[1]

5.5.1.2 Completely-Mixed Activated Sludge (Example C)

Past experience with completely-mixed domestic activated sludge on a large scale has been quite successful, although somewhat limited. Completely-mixed plants have been installed at Grand Island, Nebraska; Freeport, Illinois; South Tahoe, California; and Albany, Oregon. McKinney (13) and Smith (14) have reported the usefulness of this process for upgrading an overloaded activated sludge plant. When the completely-mixed process is to be considered as an upgrading technique for a conventional plant, the geometric configuration of the aeration basin poses a major problem. The plug flow hydraulic pattern must be altered to a completely-mixed pattern.

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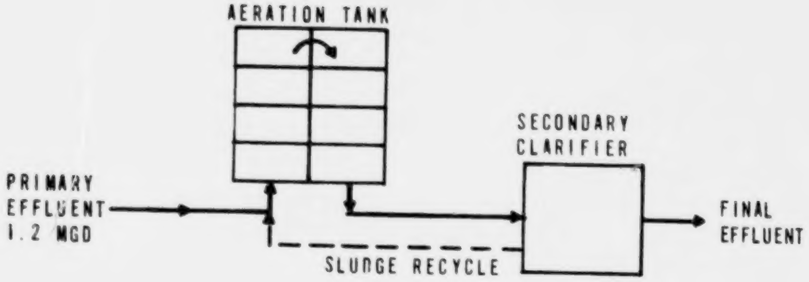
[1] These costs are based on ENR Index of 1500 and contain no contingency for engineering design, bonding, and construction supervision.

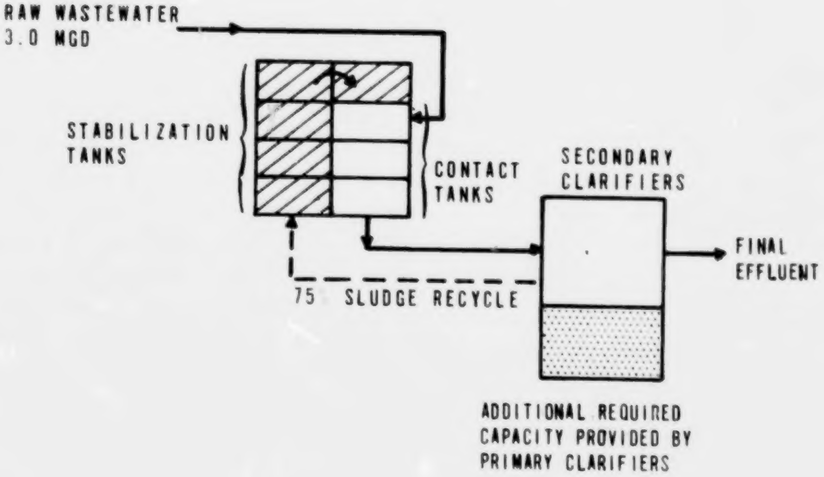
FIGURE 5-16

UPGRADING A CONVENTIONAL ACTIVATED SLUDGE PROCESS TO CONTACT STABILIZATION

EXAMPLE B


AERATION TANK
SECONDARY CLARIFIER
PRIMARY EFFLUENT 1.2 MGD
FINAL EFFLUENT
SLUDGE RECYCLE

TREATMENT SYSTEM BEFORE UPGRADING
CONVENTIONAL ACTIVATED SLUDGE (MECHANICAL AIR SYSTEM)


RAW WASTEWATER 3.0 MGD
STABILIZATION TANKS
CONTACT TANKS
SECONDARY CLARIFIERS
FINAL EFFLUENT
75 SLUDGE RECYCLE
ADDITIONAL REQUIRED CAPACITY PROVIDED BY PRIMARY CLARIFIERS

TREATMENT SYSTEM AFTER UPGRADING
CONTACT STABILIZATION PROCESS

Table 5-12

Upgrading Conventional Activated
Sludge to Contact Stabilization - Example B

Description	Overloaded Design Condition	Upgraded Design Condition
Flow, mgd	3.0	3.0
Influent BOD, mg/l	200	200
Primary Clarifier		
Overflow Rate, gpd/sq.ft.	1,200	–[1]
BOD Removal, percent	20	–[1]
Aeration Tank		
Volumetric Loading, lbs. BOD/day/1,000 cu.ft.	44	60[2]
Sludge Recycle, percent	15	75
Detention Time, hours	4.4[3]	–
Contact Basin	–	1.1[4]
Stabilization Basin	–	4.2[5]
Secondary Clarifier		
Overflow Rate, gpd/sq.ft.	960	780[6]
BOD Removal in Secondary Units	75	90
SS Removal in Secondary Units	75	90
Effluent BOD, mg/l	40	20
Effluent SS, mg/l	30	18

[1] Primary clarifier converted to secondary clarifier.
[2] Total organic loading increases due to elimination of primary treatment.
[3] Based on influent flow plus 15 percent sludge recycle to the total basin.
[4] Based on influent flow plus 75 percent sludge recycle to the contact basin.
[5] Based on 75 percent sludge recycle to the stabilization basin.
[6] Reduction in OFR is achieved by converting the primary clarifier to a secondary basin.

Example C is presented to illustrate some of the various engineering considerations which must be evaluated before implementing this type of upgrading procedure. The original flow diagram of Example C is similar to Example A shown in Figure 5-15. Original design, overloaded, and upgraded design data for Example C are presented in Table 5-13. The plant was upgraded from 5.0 mgd to 10.0 mgd using a completely mixed flow pattern as shown in Figure 5-17. The influent wastewater and recycled sludge piping were modified so that the flow would be uniformly distributed throughout the aeration tank. Four new longitudinal effluent weirs were installed in the aeration tanks to induce a traverse flow pattern as indicated in Figure 5-17. A new agitator sparger air system was installed capable of supplying 600 cu.ft. of air/lb. of BOD removed. In addition, the average sludge recycle capacity was increased to 60 percent of the upgraded flow and the clarification capacity was increased to maintain an average overflow rate of 800 gpd/sq.ft. As in Example A, an upgrading of the total plant would require consideration of expanded primary clarification and sludge handling facilities.

The capital costs for the upgrading were estimated at $700,000 ($140 per 1,000 gpd of incremental upgraded capacity) and were allocated as follows:

Aeration Tank Modifications	$280,000
Secondary Clarifier Expansion	420,000
TOTAL	$700,000[1]

5.5.1.3 Oxygen Aeration (Example D)

There are special areas for consideration in upgrading an existing conventional activated sludge plant by use of oxygen aeration. In addition to those of concern when converting to other activated sludge modifications, the following are listed:

1. The foundation or pile capacity must be checked against the increased loading of the oxygen aeration dissolution system.
2. The structural integrity of the aeration tank walls must be checked due to the increased loading, if pre-cast concrete tank covers are used.
3. Baffling may be required to sectionalize the aeration tank in order to be compatible with various oxygen aeration systems.
4. Existing tank aeration piping may have to be removed back to the tank header.

---

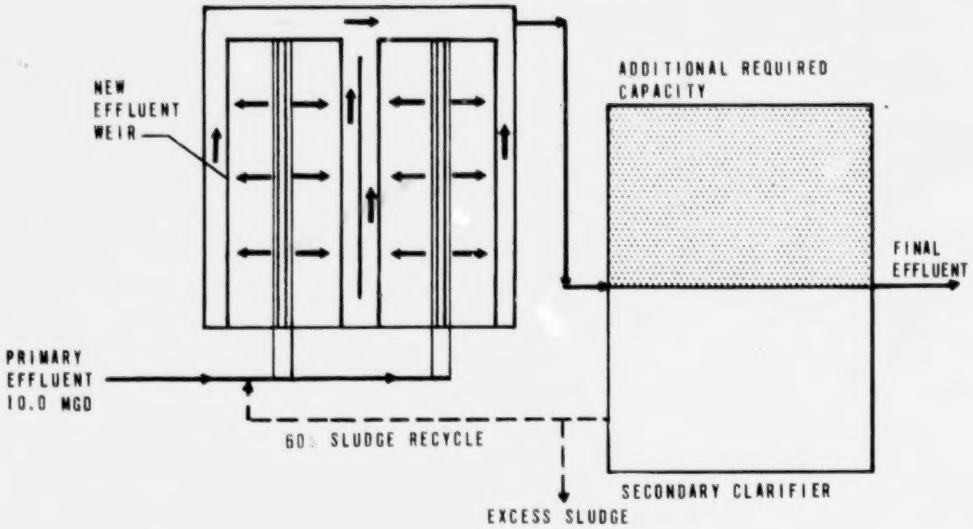
[1] These costs are based on ENR Index of 1500 and contain no contingency for engineering design, bonding, and construction supervision.

FIGURE 5-17

UPGRADING CONVENTIONAL ACTIVATED SLUDGE TO A COMPLETELY-MIXED SYSTEM

EXAMPLE C

TREATMENT SYSTEM AFTER UPGRADING TO COMPLETELY-MIXED PROCESS


NEW EFFLUENT WEIR
ADDITIONAL REQUIRED CAPACITY
FINAL EFFLUENT
PRIMARY EFFLUENT 10.0 MGD
60% SLUDGE RECYCLE
SECONDARY CLARIFIER
EXCESS SLUDGE

TYPICAL CROSS SECTION OF UPGRADED AERATION TANK (26)

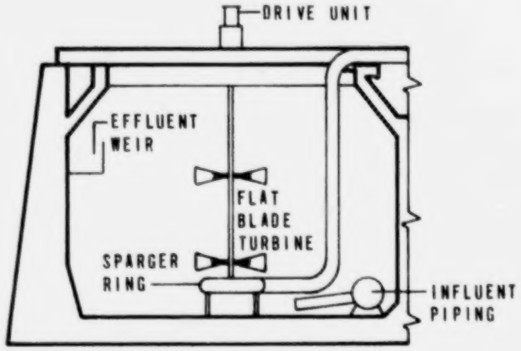

DRIVE UNIT
EFFLUENT WEIR
FLAT BLADE TURBINE
SPARGER RING
INFLUENT PIPING

Table 5-13

Upgrading Conventional Activated Sludge to a Completely-Mixed System - Example C

Description	Original Design Before Overloading	Overloaded Design Condition	Upgraded Design Condition
Flow, mgd	5.0	10.0	10.0
Influent BOD, mg/l	200	305	305
Primary Treatment			
Percent BOD Removal	30	30[1]	30
Aeration Tank			
MLSS, mg/l	2,000	2,000	3,000
Sludge Recycle, percent	25	25	60
Air Requirements, cu.ft. air/lb. BOD removed	820	--	600
Volumetric Loading, lbs. BOD/day/1,000 cu.ft.	35	107	107
Organic Loading, lbs. BOD/day/lb. MLSS	0.34	1.04	0.69
Detention Time In Aerator, minutes[2]	300	150	150
Secondary Clarifier			
Overflow Rate, gpd/sq.ft.	800	1,600	800
Secondary Treatment			
Percent BOD Removal	86	62	91
Effluent BOD, mg/l	20	80	20

[1] Requires modification of primary clarifier to handle increased hydraulic load to achieve 30 percent BOD removal.

[2] Excluding sludge recycle.

5. Provide protection against the potential explosion hazard of pure oxygen or oxygen-enriched air.

6. Provide protection against potential accelerated corrosion due to pure oxygen or oxygen-enriched air.

Some of the aspects of using oxygen aeration which may make it economically attractive include:

1. The oxygen generation equipment may be placed outside and does not require a protective enclosure.

2. Expensive renovation of the blower building is eliminated.

3. Capital and operational costs may be reduced compared to diffused air systems. However, the cost differential between the systems decreases with decreased plant size (21).

4. Reduced sludge production.

Use of oxygen aeration for upgrading municipal treatment plants handling extremely large flows is in the design stage in Detroit, Michigan, and is under consideration in New York City. In New York City, a 20-mgd section of an existing modified air aeration plant will be converted to oxygen aeration to upgrade treatment efficiency. Oxygen aeration at Detroit will be utilized to expand a 300-mgd section of the existing primary treatment plant to secondary treatment.

Example D presents the modification of a diffused air, conventional activated sludge plant to oxygen aeration. Original design data, overloaded, and upgraded design data for this example are presented in Table 5-14. The conversion to oxygen aeration permits the capacity of the plant to be increased from 2 to 6 mgd.

The capital costs include covering the existing aeration basin, oxygen generation and dissolution equipment, increasing the sludge recycle capacity to 50 percent of the upgraded flow, and maintaining a secondary clarifier overflow rate of approximately 800 gpd/sq.ft.

Table 5-14

Upgrading Conventional Activated
Sludge to an Oxygen Aeration System - Example D

Description	Original Design Before Overloading	Overloaded Design Condition	Upgraded Design Condition
Flow, mgd	2	6	6
Influent BOD, mg/l	200	200	200
Primary Treatment			
Percent BOD Removal	30	30[1]	30
Aeration Tank			
MLSS, mg/l	2,000	2,000	4,000
Sludge Recycle, percent	25	25	50
Air Requirements, cu.ft. air/lb. BOD removed	800	--	--
Oxygen Requirements, lbs. O_2/lb. BOD removed	--	--	1.2
Volumetric Loading, lbs. BOD/day/1,000 cu.ft.	35	105	105
Organic Loading, lbs. BOD/day/lb. MLSS	0.34	1.02	0.51
Detention Time in Aerator, minutes[2]	300	100	100
Secondary Clarifier			
Overflow Rate, gpd/sq.ft.	800	–	800
Secondary Treatment			
Percent BOD Removal	86	64	86
Effluent BOD, mg/l	20	50	20

[1] Requires modification of the primary clarifier to handle increased hydraulic load to achieve 30 percent BOD removal.

[2] Excluding sludge recycle.

Table 5-14

Upgrading Conventional Activated Sludge to an Oxygen Aeration System - Example D

Description	Original Design Before Overloading	Overloaded Design Condition	Upgraded Design Condition
Flow, mgd	2	6	6
Influent BOD, mg/l	200	200	200
Primary Treatment			
Percent BOD Removal	30	30[1]	30
Aeration Tank			
MLSS, mg/l	2,000	2,000	4,000
Sludge Recycle, percent	25	25	50
Air Requirements, cu.ft. air/lb. BOD removed	800	–	–
Oxygen Requirements, lbs. O_2/lb. BOD removed	–	–	1.2
Volumetric Loading, lbs. BOD/day/1,000 cu.ft.	35	105	105
Organic Loading, lbs. BOD/day/lb. MLSS	0.34	1.02	0.51
Detention Time in Aerator, minutes[2]	300	100	100
Secondary Clarifier			
Overflow Rate, gpd/sq.ft.	800	–	800
Secondary Treatment			
Percent BOD Removal	86	64	86
Effluent BOD, mg/l	20	50	20

---

[1] Requires modification of the primary clarifier to handle increased hydraulic load to achieve 30 percent BOD removal.

[2] Excluding sludge recycle.

The capital costs were estimated at approximately $700,000 ($175 per 1,000 gpd of incremental upgraded capacity) and were allocated as follows:

Aeration Tank Modifications	$130,000
Oxygen Generation and Dissolution Equipment[1]	400,000
Secondary Clarifier Expansion	170,000
TOTAL	$700,000[2]

5.5.1.4 Use of Activated Sludge Process for Treatment of Partially-Treated Effluent

This modification is by far the simplest of all upgrading procedures to implement since the activated sludge process will be built as an addition to an existing facility. The partially-treated effluent may result from a roughing filter or even an organically overloaded activated sludge process. The second-stage activated sludge process can be built using any modification as previously discussed. An economic comparison should be made before making a decision on the activated sludge modification to be used. Table 5-15 summarizes design information recommended for two-stage activated sludge when nitrification is not considered essential (33). If nitrification is required, aeration to provide at least 5 lbs. oxygen/lb. of ammonia nitrogen should be provided in addition to the air requirements for carbonaceous BOD removal.

Table 5-15

Two-Stage Activated Sludge Design Guidelines

Description	Design Parameter
Aeration Tank	
Minimum first-stage detention time, hrs.[1]	1.8
Minimum second-stage detention time, hrs.[1]	1.5
Air supply for first-stage, cu.ft. air/lb. BOD applied to plant influent	1,000
Air supply for second-stage, cu.ft. air/lb. BOD applied to plant influent	2,000
Settling Tank	
Minimum first-stage detention time, hrs.[2]	2.4
Minimum second-stage detention time, hrs.[2]	3.0
Maximum average first-stage overflow rate, gpd/sq.ft.	1,200
Maximum average second-stage overflow rate, gpd/sq.ft.	800

[1] Based on design flow not including recirculated sludge.
[2] Based on average daily design flow.

Source: Pennsylvania Department of Health (33)

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[1] Capital Costs were taken from reference (21).

[2] These costs are based on ENR index of 1500 and contain no contingency for engineering design, bonding, and construction supervision.

If year-round nitrification is a design criteria, then the following information in Table 5-16 may be useful in the preliminary sizing of the second-stage process units (34). It should be stressed that the detention time required to achieve nitrification is strongly dependent upon the temperature of wastewater and the concentration of mixed liquor solids maintained in the system.

Table 5-16

Design Guidelines for Second-Stage Units
to Include Consideration of Nitrification

Description	Design Parameter
Aeration Tank	
Optimum pH range	8.2 to 8.6
Maximum Influent BOD, mg/l	40 to 50
Tank Configuration	plug flow
MLVSS, mg/l	1,000 to 2,500
D.O. at average loading, mg/l	3.0
Minimum D.O. at peak loads, mg/l	1.0
Sludge recirculation, percent	50 to 100
Detention time based on average flow, hr.	2 to 6
Oxygen requirements (stoichiometric), lbs. O_2/lb. NH_3-N	4.6
Settling tank	
Average allowable overflow rate, gpd/sq.ft.	800

5.5.2 Upgrading to Increase Organic Removal Efficiency

Upgrading techniques previously discussed relate to the ability of existing facilities to handle increased hydraulic or organic loads by providing modifications to meet existing effluent standards. However, there may be a need to meet higher effluent standards even though the existing facilities are not hydraulically or organically overloaded. Table 5-17 contains suggested alternatives for improving effluent quality under these conditions. The main purpose of the table is to present various alternatives and to suggest a range of anticipated improvement in performance for each alternative.

It should be emphasized that, in cases where unit processes are added on to existing facilities, the improvement in overall organic removal will be a direct function of the BOD removal achieved in the "add-on" process. However, where unit processes precede existing units, e.g. the use of a roughing filter, the overall BOD removal may not be increased in direct proportion to the amount achieved by the "add-on" process.

Table 5-17

Upgrading Techniques for Improvement of Activated Sludge Treatment Plant Efficiency

Addition Preceding Existing Unit	Existing Process	Addition Following Existing Unit	Incremental BOD Removal Across the Added Process percent
	Activated Sludge		
Roughing Trickling Filter (Rock or Synthetic Media)			20-40
Chemical Addition To Primary Clarifier			30-50
		2nd Stage Activated Sludge[1]	30-70
		Polishing Lagoon	30-60
		Multi-media Filters	50-80
		Microstraining	30-80
		Activated Carbon	60-80

[1] A consideration if year-round nitrification is required.

A detailed discussion on polishing lagoons, microstrainers, filters, activated carbon, and clarifier modifications appears in subsequent chapters. The applicability of these alternatives to individual cases should be evaluated in detail prior to the implementation of a particular upgrading procedure.

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CHAPTER 6

CLARIFICATION AND CHEMICAL TREATMENT

6.1 General

Improved solids separation in primary and secondary clarifiers, either by operational changes or by addition of chemicals, is usually accompanied by concurrent reductions in BOD values in the overflow. Therefore, improvements in the clarification process can be advantageously used to meet specific treatment requirements, particularly where the treatment plants are experiencing hydraulic and organic overloads. The solids separation process can be improved by adding additional clarification area, by chemical treatment of wastewaters, or by use of more efficient settling devices.

6.2 Primary Clarification

Increased solids separation in primary clarifiers has the following advantages in addition to increasing their hydraulic capacity:

1. An increase in quantity of primary sludge produced (which can be more readily thickened and dewatered than secondary sludge).

2. A decrease in quantity of secondary sludge produced.

3. A decrease in organic loading to secondary treatment process units.

The primary clarifier performance significantly influences the extent of secondary treatment required and, in most cases, affects the overall effluent quality of existing treatment plants. Also, since clarification is the most economical way to remove suspended and colloidal pollutants, every effort should be made to improve the primary clarification process before additional facilities are considered.

6.3 Secondary Clarification

The performance of conventional secondary wastewater treatment systems is determined by comparing the quality of the overflow from secondary clarifiers to that of the incoming wastewater. The biological treatment unit converts a portion of the soluble and insoluble organic pollutants to suspended organic solids (biological). Unless these organic solids are effectively removed in the secondary clarifiers, the treatment process cannot be considered a successful operation. Fortunately, the biological solids flocculate and separate readily by gravity, provided that the rise velocity in the clarifier is maintained below the settling velocity of the floc particles. The following conditions, either alone or in combination, will disrupt the secondary clarifier performance:

1. Hydraulic overloading, which causes the rise velocity of the wastewater in the secondary clarifier to exceed the settling velocity of the solids.

2. Organic overloading of the biological treatment units, which results in an increased solids load to secondary clarifiers.

In addition to the above conditions, improper inlet and outlet design often cause short circuiting of wastewater, thereby reducing the overall BOD removal efficiency. Improper sludge withdrawal techniques can also cause solids carryover to the clarifier effluent.

6.4 Chemical Treatment

Chemical addition to primary and secondary clarifiers in this manual is concerned only with increased solids and BOD removal. Chemical treatment for phosphorus removal is covered in detail in the process design manual for phosphorus removal.

6.4.1 Chemical Addition to Improve Clarification

At first, chemicals were used to improve the efficiency of primary clarification systems. Later, when these systems were followed by secondary treatment processes, the practice of adding chemicals to upgrade primary treatment because unnecessary. However, the technique of adding chemicals to the primary clarifier is still an effective upgrading procedure for a secondary plant when the following conditions exist (1):

1. Wastewater flow is intermittent or varies greatly.
2. Space available for additional clarification facilities is limited.
3. Industrial wastes that would interfere with biological treatment are present.
4. Plant is hydraulically and/or organically overloaded.
5. Improvements in existing treatment performance are required as an interim measure before the addition of new facilities.

The chemicals commonly used in wastewater treatment are the salts of iron and aluminum lime, and synthetic organic polyelectrolytes. The iron (ferrous and ferric) and aluminum salts (sodium aluminate or alum) react with the alkalinity and soluble orthophosphate in wastewater to form precipitates of the respective metallic hydroxides or phosphates. In addition, they destabilize the colloidal particles that would otherwise remain in suspension. These precipitates, along with the destabilized colloids, flocculate and settle readily in a clarifier.

Sodium aluminate is a basic salt and can be advantageously used for wastewaters containing low to moderate amounts of alkalinity. Alum, being an acid salt, is best suited for wastewater high in alkalinity. While both alum and sodium aluminate exhibit great capability for total phosphorus removal, the use of alum introduces six times as much dissolved solids to the wastewater as does sodium aluminate (2). Normally, lime is used to precipitate hydrous oxides of iron and aluminum when the alkalinity of wastewaters is low. The reaction of iron and aluminum salts is pH-dependent and has to be evaluated for each case to determine the most effective pH range and the optimum chemical dosage.

The addition of lime alone is also effective in coagulating wastewater. The positive calcium ions help to destabilize colloidal particles while precipitating soluble orthophosphates as hydroxyapatite. Since lime treatment takes place at high pH (9.0-11.5), the effluent from this process will normally require pH adjustment before biological treatment. In some cases, natural recarbonation from biological oxidation is adequate to maintain the pH within acceptable limits.

6.4.2 Use of Chemicals in Primary Clarifiers

The effect of polyelectrolyte addition (used either alone or in combination with inorganic coagulants) on primary clarifier performance is shown in Table 6-1. For comparative purposes, the performance of the clarifiers before and after the addition of chemicals is shown. As indicated, the average values for suspended solids and BOD removals were 37.7 percent and 31 percent, respectively, before chemical treatment. As a result of chemical addition, suspended solids and BOD removal efficiencies increased to 64.7 percent and 46.7 percent, respectively. It is also evident from Table 6-1 that the effect of polyelectrolyte addition was pronounced where the existing clarifier performance was poor, as indicated by initial low suspended solids removal. The above data illustrate that the proper selection and application of polyelectrolytes and chemicals to raw wastewater can significantly improve primary clarifier performance.

When considering the addition of chemicals to primary clarifiers, it is important to examine the effect of increased primary clarifier efficiency on subsequent treatment units. The increased removal of suspended solids and BOD from raw wastewater can affect the downstream biological process in several ways. If the BOD load to the aerator falls below 0.25-0.35 lb. BOD/lb. MLVSS/day for extended periods of time, nitrification conditions can develop in the aerator. This can reduce the total oxygen demand of the effluent, but will impose an added oxygen demand on the aeration facility because the oxidation of one pound of ammonia nitrogen requires about 4.5 pounds of oxygen.

A decrease in loading to the aerator will normally require more careful management of sludge to insure stable operation of the aeration basin. However, the quantity of excess activated sludge generated under these reduced loading conditions will be substantially less than that generated under normal loading conditions, and this may be considered an added advantage of adding chemicals to the primary clarifier.

Little information has been generated regarding the periodic addition of chemicals to the primary clarifier for controlling peak organic or hydraulic loads. This approach, while not always applicable, can frequently be used to maintain system stability during temporary overload.

Table 6-1

Effect of Chemical Treatment on Primary Clarifier Performance

Type and Amount of Chemical Added	Performance Preceding Chemical Treatment: SS Removed mg/l	SS Removed percent	BOD Removed mg/l	BOD Removed percent	Weight Ratio of WAS/PS[1]	Performance After Chemical Treatment: SS Removed mg/l	SS Removed percent	BOD Removed mg/l	BOD Removed percent	Weight Ratio of WAS/PS[1]	Reference
Purifloc - A21 (0.95 mg/l)	13	12	28	26	0.61	75	65	46	48	0.31	3
DOW - SA1193 (0.2 mg/l)	13	12	28	26	0.61	72	55	36	37	0.41	3
Purifloc - A21 (1 mg/l)	157	43	82	23	—	281	76	127	33	—	3
Purifloc - A21 (0.75 mg/l)	26	18	—	—	—	69	52	—	—	—	3
Purifloc - A21 (0.89 mg/l)	113	43	50	22	—	159	60	87	37	—	3
DOW - SA1193 (0.25 mg/l)	120	47	—	—	0.8	151	61	—	—	0.46	3
Purifloc - A21 (1 mg/l)	107	47	135	37	—	169	62	154	46	—	3
$FeCl_2$ + NaOH + Purifloc - A23 (0.3 mg/l)	230	62	111	31	—	379	79	74	39	—	4,5
$FeCl_2$ + NaOH + Purifloc - A23 (0.3 mg/l)	104	49.7	83	43.8	—	173	76.8	105	57.8	—	4,5
Purifloc - A21 (1 mg/l)	—	—	—	—	0.79	—	—	—	—	0.28	6
Purifloc - A23 (0.25 mg/l)	52	31	47	31	1.44	80	51	58	46.4	0.67	7
$FeCl_3$ + Purifloc - A23	93	33	53	34	—	196	74	102	61	—	8
$FeCl_3$ + Purifloc - A23	93	33	53	34	—	213	68	97	53	—	8
Purifloc - A21 (0.74 mg/l)	—	50	—	36	—	—	63	—	45	—	9
Purifloc - A21M (1.14 mg/l)	—	43	—	—	—	—	63	—	—	—	9
MEAN	93.5	37.7	68.5	31	—	168	64.7	87	46.7	—	

[1] WAS - Waste activated sludge
PS - Primary sludge

Schmidt and McKinney (10) studied phosphorus removal by lime addition to the primary clarifier of a treatment system which also included secondary treatment. In this study, the system was operated at a pH value of 9.5, which during biological treatment was reduced to a value between 7 and 8. Therefore, no neutralization was required. The lime precipitation step reduced the BOD by 60 percent, suspended solids by 90 percent, and total phosphorus by 80 percent. However, Schmidt and McKinney indicated that the lime-primary sludge was gelatinous in nature and required polyelectrolyte treatment prior to dewatering by vacuum filtration. They further indicated that the mass of primary sludge is about twice that obtained by conventional settling, although the total mass of primary and secondary sludge produced is increased by less than 50 percent (10). Lime addition to primary clarifiers for phosphorus removal has been used in many locations. In all cases, significant improvements in both suspended solids and BOD removal were noted. Table 6-2 presents the results of some of these studies.

Table 6-2

Lime Addition to Primary Clarifiers

Location	Lime Added mg/l CaO	Percent Removal Before Lime Addition		Percent Removal After Lime Addition		Remarks	Reference
		BOD	SS	BOD	SS		
Duluth, Minnesota	75	50	70	60	75	----	12
	125	55	70	75	90	----	12
Rochester, New York	140	—	—	50	80-90	Jar tests	12
Lebanon, Ohio	145	—	—	66	74	Pilot plant	11

As mentioned above, the addition of lime to the primary clarifier can be expected to increase the primary sludge mass to about twice that obtained by conventional primary settling, depending on the alkalinity of the incoming wastewater. Therefore, a complete evaluation of the sludge handling facilities must be made when considering this technique. For instance, some states have cautioned against this practice when the primary sludge is to be anaerobically digested.

Freese, et al, (9) studied the application of polyelectrolytes for raw wastewater flocculation in the District of Columbia's Water Pollution Control Plant. The plant also recirculated thickener overflow and digested sludge elutriate to the primary clarifier. Even though the primary clarifier performance improved with the addition of chemicals, the solids input from the sludge elutriation process remained the same. As a result, the full benefit of polyelectrolyte addition was not realized. This indicated that separate treatment of elutriate is required, since polyelectrolyte addition apparently does not enhance the capture of the fine solids which are normally discharged in the elutriate. These fine solids often accumulate in the solids-handling system

Mogelnicki (13) reported on the effect of polyelectrolyte addition in primary clarifiers on the overall BOD removal. The data reported by Mogelnicki, covering both activated sludge and trickling filter processes, are shown in Table 6-3. These data indicated that polyelectrolyte addition to primary clarifiers increases the overall BOD removal by approximately 7 percent.

Table 6-3

Effect of Polyelectrolyte Addition in Primary Clarifier on Overall BOD Removal

Type of Wastewater	Primary Clarifier				Total Plant			
	Percent Removal Before Polyelectrolyte Addition		Percent Removal After Polyelectrolyte Addition		Percent Removal Before Polyelectrolyte Addition		Percent Removal After Polyelectrolyte Addition	
	BOD	SS	BOD	SS	BOD	SS	BOD	SS
Activated Sludge[1]	26	—	48	—	83	—	90	—
Trickling Filter[1]	23	43	33	76	79	72	85	84

[1] With 1 mg/l Purifloc A-21

Source: Mogelnicki (13)

6.4.3 Use of Chemicals in Secondary Clarifiers

There is little published information available on the use of chemicals to improve secondary clarifier performance. This is probably due to the fact that aerobic biological sludges flocculate and settle readily if normal growth conditions are maintained. However, upsets in secondary clarifier performance can occur as a result of increased hydraulic and/or solids loading or development of a filamentous or bulking sludge. When one or more of these conditions exist, the use of inorganic chemicals and/or polyelectrolytes has been successful in some instances in obtaining a satisfactory effluent.

Singer, et al. (14) studied the effect of adding cationic and anionic polyelectrolytes to improve settling characteristics of bulking activated sludge in the laboratory. Their studies indicated that cationic polyelectrolytes at a concentration of 2-3 mg/l were effective in coagulating a bulking activated sludge but that the anionic polyelectrolyte tested had no effect on improving settling. Goodman and Mikkelson (15), on the basis of full-scale studies, concluded that application of cationic polyelectrolytes to primary clarifier effluent at the rate of 0.1 lb./ton of secondary dry solids increased overall BOD removal efficiency to 95 percent and decreased the loss of solids in the secondary effluent of the activated sludge plant.

Based on studies conducted at the Hanover treatment plant by the Metropolitan Sanitary District of Chicago, Zenz and Pivnicka (16) have shown that the addition of alum to aeration tanks (primarily intended for soluble phosphorus removal) improved flocculation of activated sludge. However, their results indicated that increasing amounts of alum floc escaped through the final clarifiers as the dosage of alum increased from an Al:P weight ratio of 1.54 to 1.85. The addition of alum to the aeration tank favored the development of lower organisms, while the higher forms such as protozoa and metazoa were absent when using alum. Zenz and Pivnicka also indicated that the precipitated phosphate was not released during anaerobic digestion and was permanently removed from the treatment system.

Laughlin (17) has reported adding 460 gallons/day of alum (17 percent alum solution) to the secondary clarifier of a trickling filter plant treating 1.5 mgd of wastewater. The preliminary results indicated a reduction in effluent BOD and SS concentrations from 20 mg/l and 15 mg/l, respectively, to 10 mg/l and 10 mg/l. The phosphorus concentration in the effluent was reduced from 8 mg/l to 1 mg/l. In addition, Laughlin (17) has reported problems of reduced alkalinity in sludge undergoing digestion when alum was used as coagulant in primary treatment at the Richardson, Texas plant. The addition of alum to raw wastewater was discontinued after 9 days total operation to prevent pH depression in the digester.

From the above studies, it can be concluded that alum, iron, or polyelectrolyte addition, either in the primary or secondary treatment process, can be used advantageously to improve the overall performance of the treatment system including phosphorus removal. Lime addition may not be feasible for upgrading activated sludge secondary clarifiers because of the potential adverse effect of recirculated lime sludge on mixed liquor microbial characteristics. Lime addition to either trickling filter or activated sludge secondary clarifiers will require pH adjustment of the effluent before discharge to the receiving waters. Lime addition to primary clarifiers may be used, if consideration is given to controlling the pH within acceptable limits for the subsequent processes, and to changes in sludge characteristics and handling requirements.

6.5 Other Approaches to Improvement of Clarification

6.5.1 Design and Operational Factors

In many cases, poor clarifier performance is the result of poor operation or inadequate design even when the hydraulic load has not exceeded the design values. It is essential to correct these deficiencies through modifications before any consideration is given to other upgrading techniques, such as chemical addition. Inadequate design factors which affect clarifier performance include the following:

1. Poor inlet and outlet design.
2. Poor sludge withdrawal system.
3. Absence of scum removal devices.

Poor inlet or outlet design can cause excessive turbulence or short circuiting in the clarifiers, resulting in the escape of solids in the effluent. Fall (18) has described a system in the Greater Peoria Sanitary District Sewage Treatment Plant (employing the Kraus Modification) where changes in inlet and outlet design improved clarifier performance. He reported that conversion of center-feed square primary and secondary clarifiers to peripheral-feed systems permitted overflow rates as high as 4,100 gpd/sq.ft. without any apparent loss in solids removal efficiency, and further that combinations of primary and secondary sludge were concentrated to as much as 6 percent in the clarifiers. Typical cross sections of a circular clarifier with center and peripheral feeds are illustrated in Figure 6-1.

Poor sludge withdrawal systems in secondary clarifiers often cause sludge accumulation, thus creating oxygen-deficient conditions. As a result, the sludge will sometimes gasify due to denitrification, and will rise to the surface and overflow the weirs. This condition can be partially corrected by installing automatic sludge-withdrawal devices, sludge blanket finders, and sludge density meters to facilitate proper sludge management practices. Instrumentation commonly used for this purpose is discussed in detail in Chapter 14.

West (19) has described several case histories where improved operational conditions increased the efficiency of clarifiers. Table 6-4 shows the methods and performance obtained from the above studies. Installation of a suction-type sludge-withdrawal device, in lieu of a scraper mechanism, and of a scum removal device is strongly recommended for improving secondary clarifier performance. The scraper mechanism and hydraulic suction device for sludge removal are illustrated in Figure 6-1.

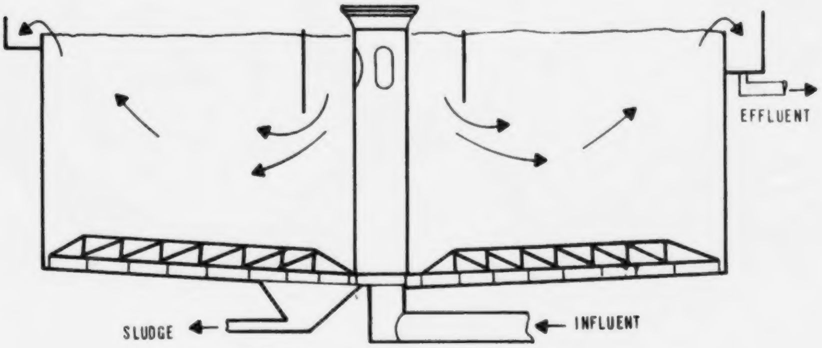
6.5.2 Use of Tube Settlers

According to the classical theory of discrete particle settling, the efficiency of suspended particle removal in a sedimentation basin is solely a function of overflow rate and is independent of depth and detention time. If the above theory is applicable to raw wastewater or activated sludge floc settling, then the clarifier performance could be improved by introducing a number of trays or tubes in the existing clarifiers. However, the introduction of trays has been found to be unsuitable on a practical basis because of the sludge collection and removal devices required. A new device called a tube settler, which employs the above approach of trays or tubes in the existing clarifiers, has recently become available. Tube settlers (of various lengths) are usually installed in modules at an angle to the horizontal, and are made of a light-weight, durable material. A typical inclined-tube settler module is shown in Figure 6-2 (20). Figure 6-3 shows a typical clarifier cross-section with the tube settler modules installed (20). Even though tube settlers can be used at any angle of inclination, Hernandez and Wright (21) have recommended an angle of 60°, for self-cleaning purposes.

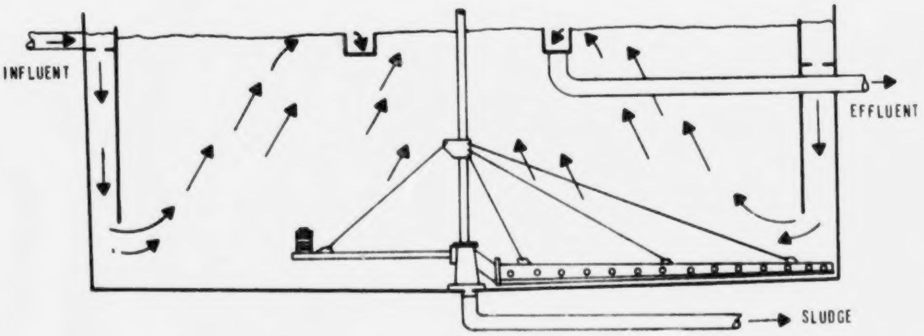
Tube settlers have been used in primary and secondary clarifiers to improve performance as well as to increase throughput in existing clarifiers. Conley and Slechta (22) and Culp, et al (20) have described the performance of several plant-scale installations of tube settlers in primary and secondary clarifiers. The results of their studies indicate that the overflow rates in primary clarifiers can be increased to 5,000 gpd/sq.ft. while producing the same quality effluent as the control unit without the settlers. Tube settlers enhance the ability to capture settleable solids at high overflow rates because the depth of settling has been reduced to a few inches in the tube. It should be realized that tube settlers do not improve

FIGURE 6-1

TYPICAL CLARIFIER FEED AND SLUDGE REMOVAL MECHANISMS


EFFLUENT
SLUDGE
INFLUENT

CIRCULAR CENTER FEED CLARIFIER WITH
A SCRAPER SLUDGE REMOVAL SYSTEM


INFLUENT
EFFLUENT
SLUDGE

CIRCULAR PERIPHERAL FEED CLARIFIER WITH A
HYDRAULIC SUCTION SLUDGE REMOVAL SYSTEM

Table 6-4

Effect of Clarifier Operational Improvements on Overall Effluent Quality

Plant Location & Capacity	Operational Improvement	Improvement in Overall Effluent Quality
Sioux Falls, South Dakota	Sludge Blanket Finder for Secondary Clarifier	Effluent BOD reduced from 20 to 10 mg/l
3.5 mgd	Turbidimeter for Effluent Quality	Effluent SS reduced from 35 to 13 mg/l
	Increased Sludge Recirculation Capacity	
Metropolitan St. Louis Sewer District, Missouri	Sludge Blanket Finder for Secondary Clarifier	Effluent BOD reduced from 40 to 9 mg/l
21 mgd	Turbidimeter for Effluent Quality	Effluent SS reduced from 92 to 16 mg/l
	Increased Air Supply	BOD removal efficiency increased from 73 to 94 percent
	Reduced Sludge Recirculation	SS removal efficiency increased from 46 to 92 percent

Source: West (19)

FIGURE 6-2

INCLINED TUBE SETTLER MODULE (20)

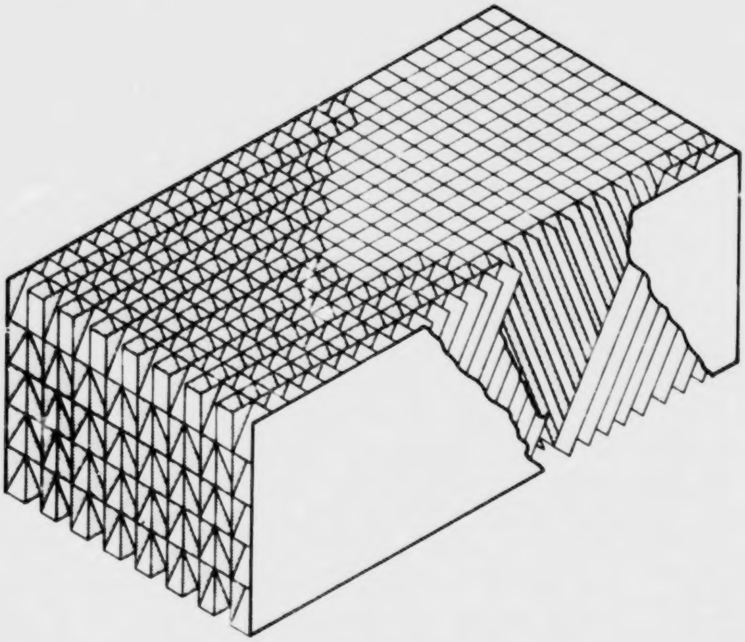
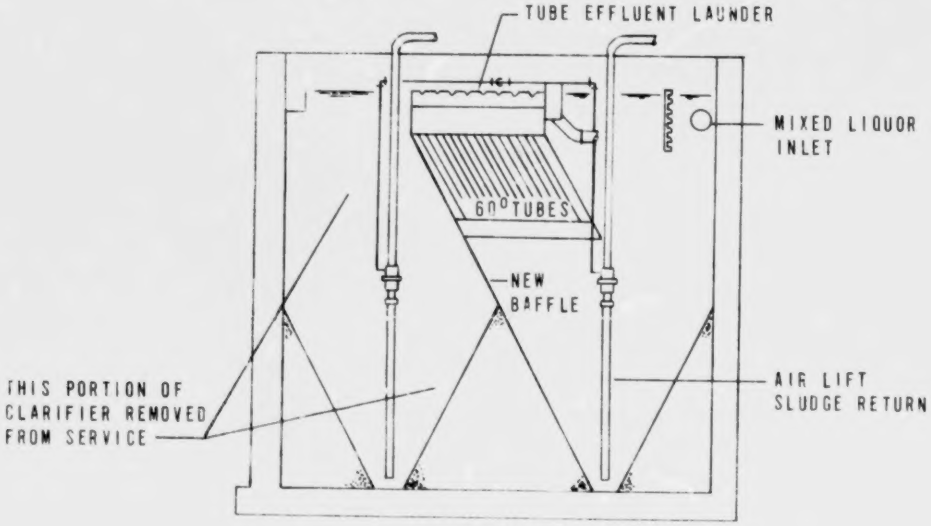


FIGURE 6-3

INSTALLATION OF TUBE SETTLERS AT THE WICKAM, PENNSYLVANIA SEWAGE TREATMENT PLANT (20)


TUBE EFFLUENT LAUNDER
MIXED LIQUOR INLET
60° TUBES
NEW BAFFLE
THIS PORTION OF CLARIFIER REMOVED FROM SERVICE
AIR LIFT SLUDGE RETURN

the efficiency of primary clarifiers that are already achieving very high (40-60 percent) removals of suspended solids. Moreover, tube settlers will neither remove colloidal solids that remain in suspension nor induce additional coagulation to effect added particle removal.

Tube settlers have been used to improve secondary clarifier performance where the clarifiers were subjected to overflow rates of 900 to 2,800 gpd/sq.ft. (20) (22). Prior to the installation of tube settlers, the effluent solids concentration varied between 8 and 1,480 mg/l. This range was reduced to 4-156 mg/l after the installation.

Fouling due to attachment and growth of biological slime on the sides of the tubes is sometimes a problem. Some form of cleaning device (water jet or air) is required so that the solids build-up can be removed occasionally. Conley (22) has recommended the use of 1.0 gpm/sq.ft. as a maximum overflow rate and 35 lbs./sq.ft./day as a maximum solids loading for the design of secondary clarifiers with tube settlers. Since the flocculating and settling characteristics of sludge vary from plant to plant, each case should be evaluated separately for suitable design criteria. Small pilot units are available from the manufacturer for this purpose.

The performance of clarifiers provided with tube settlers at various installations is summarized in Table 6-5. Little information is available at the present time to establish cost information on tube settlers. However, an estimating cost figure of 12 to 20 dollars/sq.ft. for tube settlers with an installation cost of 5 to 15 dollars/sq.ft. has been recommended by the manufacturer (22).

6.6 Chemical Feeders

Table 6-6 contains a summary of properties and characteristics of chemicals commonly used in wastewater treatment. Most chemicals used in wastewater treatment are added to the unit treatment process in solution. Dry chemicals may be fed to dissolving tanks by either volumetric of gravimetric feeders. Gravimetric feeders are more accurate and dependable, but cost more than volumetric feeders.

One type of volumetric feeder uses a continuous belt from under the hopper to the dissolving tank. A mechanical gate mechanism regulates the depth of material on the belt, and the rate of feed is governed by the speed of the belt and/or the height of the gate opening. The hopper normally is equipped with a vibratory mechanism to reduce arching. This type of feeder is not usually suited for easily fluidized materials. Another type employs a screw or helix. Rate of feed is governed by the speed of screw or helix rotation. Some screw-type designs are self-cleaning, while others are subject to clogging.

Most of the other types of volumetric feeders fall into the positive-displacement category, involving some form of moving cavity of a specific or variable size. In operation, the chemical falls by gravity into the cavity and is more or less fully enclosed and separated

Table 6-5

Performance of Clarifiers Using Tube Settlers

Plant Location	Type	Size mgd	Tube Location	Existing Facility SOR[1] gpm/sq.ft.	Existing Facility Eff. SS mg/l	Operational Data Using Tube Settlers SOR[1] gpm/sq.ft.	Operational Data Using Tube Settlers Eff. SS mg/l
Philomath, Oregon	Trickling Filter	0.15	Secondary Clarifier	0.6	60-70	3.3-4.6	60-70
Philomath, Oregon	Trickling Filter	0.15	Primary Clarifier	0.84	40-45[2]	2.1-3.3	34-41[2]
Hopewell Township, Pennsylvania	Activated Sludge	0.13	Secondary Clarifier	0.34	60-70	2-3	27
Miami, Florida	Activated Sludge	1.0	Secondary Clarifier	1.3	500	1.7	33

[1] SOR - Surface Overflow Rate.
[2] Percent removal rather than concentration.

Source: Conley and Slechta (22)

Table 6-6

Properties and Characteristerics of Selective Chemicals Used in Wastewater Treatment

Chemical Name and Formula	Shipping Container	Weight lbs./cu ft	Solubility in Water gm./100 cc.	Storage Container Materials	Handling Characteristics	Feed Regulation	Strength of Solution (%) and Characteristics	Suitable Handling Material for Solution
Aluminum sulfate $Al_2(SO_4)_3 \cdot 14H_2O$	100-200 lbs Bags or Bulk	60-67	78.8 (30°C)	Iron, Steel	Dusty	Solution	<6% Acid and Corrosive	Lead, Rubber, Plastics
Ferric Chloride $FeCl_3$ Solution	Barrels, Bulk	12 lbs./gal.	35-45% Solution	Glass, Rubber, Concrete	Acid, Corrosive	—	<45%	Plastic, Glass
$FeCl_3 \cdot 6H_2O$	Barrels	60-65	91.1 (20°C)	Glass, Rubber, Concrete	Acid, Corrosive	Solution	Acid and Corrosive	Rubber
$FeCl_3$ - Anhydrous	Drums	65-70	74.4 (0°C)	Glass, Rubber, Concrete	Acid, Corrosive	Solution	Acid and Corrosive	—
Ferric Sulfate $Fe_2(SO_4)_3 \cdot 9H_2O$	Bags, Drums	70-72	Very Soluble	Plastic, Rubber, Stainless Steel	—	Solution	<25% Acid and Corrosive	Plastic, Glass, Rubber
Ferrous Sulfate $FeSO_4 \cdot 7H_2O$	Bags, Barrels, Bulk	63-66	60.2 (30°C)	Asphalt, Concrete	—	Solution	<6% Acid and Corrosive	Plastic, Glass
Sodium Aluminate[1] $Na_2O \cdot Al_2O_3$	Bags, Drums	50-60	—	Iron, Plastic, Rubber	Need Hopper Agitation	Solution	—	Iron, Plastic, Rubber
Calcium Hydroxide $Ca(OH)_2$	50 lbs. Bags Barrels, Bulk	25-70	23.4 (40°C)	Asphalt, Concrete, Rubber	Dusty	Slurry	<10% Alkali	Plastic, Rubber
Calcium Oxide CaO	50 lbs. Bags, Barrels, Bulk	55-70	Forms $Ca(OH)_2$	Asphalt, Concrete, Rubber	Dusty	Slurry	<25% Alkali	Plastic, Rubber
Polyelectrolyte[2]	50 lbs. bags	75	—	—	—	Solution	0.25-6.0	—

[1] Also available in liquid form

[2] There are over 200 polyelectrolytes marketed for wastewater treatment. Information on particular polyelectrolyte is available from the manufacturer

Sources: Water Pollution Control Federation (1)
Fair and Geyer (23)

from the hopper's feed. The size of the cavity and the rate at which the cavity moves and is discharged govern the amount of material fed. The positive control of the chemical may place a low limit on rates of feed. One unique design is the progressive-cavity metering pump, a non-reciprocating type. Positive-displacement feeders often utilize air injection to enhance flowability of the material.

The basic drawback of volumetric feeder design, i.e. its inability to compensate for changes in materials density, is overcome by modifying the volumetric design to include a gravimetric or loss-in-weight controller. This modification allows for weighing of the material as it is fed. The beam balance type measures the actual mass of material, and is considerably more accurate over a period of time than the less common spring-loaded gravimetric designs.

Gravimetric feeders are used where feed accuracy of about 99 percent is required for economy, as in large-scale operations, and for materials which are used in small, precise quantities. It should be noted, however, that even gravimetric feeders cannot compensate for weight added to the chemical by excess moisture. Many volumetric feeders may be converted to a loss-in-weight basis by placing the entire feeder on a platform scale tared to neutralize the weight of the feeder.

Good housekeeping and need for accurate feed rates dictate that the gravimetric feeder be shut down and thoroughly cleaned on a regular basis. Although many of these feeders have automatic or semi-automatic devices which compensate to some degree for accumulated solids on the weighing mechanism, accuracy is affected, particularly on humid days, when hygroscopic materials are fed. In some cases, built-up chemicals can actually jam the equipment.

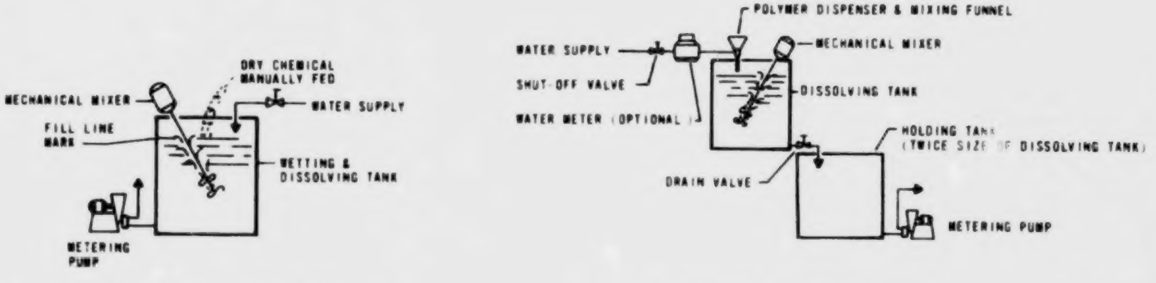
No discussion of feeders is complete without at least a passing reference to dissolvers, because any metered material must be accurately mixed with water to provide a chemical solution of desired strength. Most feeders, regardless of type, discharge their material to a small dissolving tank, which generally is equipped with a nozzle system and/or mechanical agitator depending on the solubility of the chemical being fed.

One particular area that requires careful consideration is the dispersion of dry polyelectrolytes to make feeding solutions. Long-chain polymers are very difficult to dissolve, and special equipment is often necessary. Figure 6-4 depicts three typical techniques for dissolving polymers (24). The simplest method is the manual-vortex technique, in which dry polymer is manually dropped into the vortex produced by the mixer. The manual-aspiration technique utilizes the aspiration principle to wet the polymer; water at a pressure of at least 20 psi is forced into the mixing funnel producing a downward stream which traps the dry polymer.

The third method illustrated in Figure 6-4 is the automatic-wetting spray technique. This system tends to replace the manual measuring of chemicals and free the plant operator

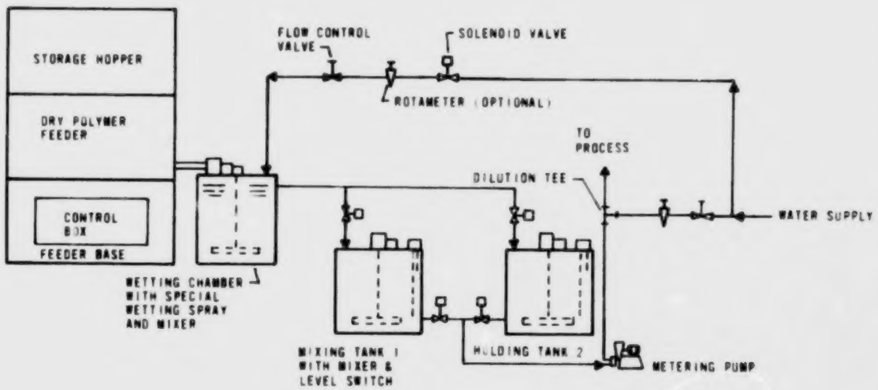
FIGURE 6-4

TYPICAL HANDLING AND APPLICATION TECHNIQUES FOR POLYELECTROLYTES (24)


MECHANICAL MIXER
DRY CHEMICAL MANUALLY FED
WATER SUPPLY
FILL LINE MARK
WETTING & DISSOLVING TANK
METERING PUMP
POLYMER DISPENSER & MIXING FUNNEL
WATER SUPPLY
MECHANICAL MIXER
SHUT-OFF VALVE
DISSOLVING TANK
WATER METER (OPTIONAL)
HOLDING TANK (TWICE SIZE OF DISSOLVING TANK)
DRAIN VALVE
METERING PUMP

A. MANUAL-VORTEX TECHNIQUE

B. MANUAL-ASPIRATOR TECHNIQUE


STORAGE HOPPER
DRY POLYMER FEEDER
CONTROL BOX
FEEDER BASE
FLOW CONTROL VALVE
SOLENOID VALVE
ROTAMETER (OPTIONAL)
TO PROCESS
DILUTION TEE
WATER SUPPLY
WETTING CHAMBER WITH SPECIAL WETTING SPRAY AND MIXER
MIXING TANK 1 WITH MIXER & LEVEL SWITCH
HOLDING TANK 2
METERING PUMP

C. AUTOMATIC-WETTING SPRAY TECHNIQUE

for other duties. Polymer is discharged evenly onto a water spray in a small wetting chamber, where it is trapped and dropped into the vortex. The entire operation may be controlled automatically by level switches and solenoid valves. Once the polymer is in solution, no agitation is required. However, some manufacturers recommended diluting the solution further before applying it to the unit treatment process.

Positive displacement or plunger-type pumps generally are used to meter polymer solutions, but diaphragm pumps can also be used and/or adapted to handle them For the more viscous solutions (>1,000 cp), the speed of a mechanical diaphragm pump should be limited to 105 strokes/minute (24). In addition, ball-type suction and discharge valves are recommended. The design of polymer-dispersing systems should recognize the temperature dependence of viscosity and the consequent effects on measuring and pumping.

The capacity of a dissolver is based on detention time, which is directly related to the solubility or wettability of the chemical. Therefore, the dissolver must be large enough to provide the necessary detention for both the chemical and the water at the maximum feed rate. At lower feed rates, the strength of solution or suspension leaving the dissolver will be less, but the detention time will be approximately the same unless the water supply to the dissolver is reduced. When the water supply to any dissolver is controlled for the purpose of forming a constant-strength solution, then mixing within the dissolver must be accomplished by mechanical means, because sufficient power will not be available from the mixing jets at low rates of flow.

Specific factors influencing chemical feed rates per volume of water, detention times, and selection of materials of construction are available in the literature (25). Alum, lime, and ferrous sulfate have been found to require about 5 minutes detention time at about 0.5 lb./gallon. Ferric sulfate requires longer detention times (20 to 30 minutes) than the other granular chemicals. Further practical experience with a number of these chemicals is available in Culp and Culp (26). Hot-water dissolvers decrease the required tank volume.

The foregoing descriptions give some indication of the wide variety of materials involved. Because of this variety, a modern facility may contain any number and variety of feeders, with combined or multiple materials capability. Ancillary equipment to the feeder also varies according to the material to be handled. Liquid feeders involve a limited number of design principles, principally to account for density and viscosity ranges. Solids feeders, relatively speaking, vary considerably due to the wide ranges of physical and chemical characteristics, feed rates, and the degree of precision and repeatability required.

6.7 Process Designs and Cost Estimates

Process units were designed and capital costs were developed for three examples of chemical treatment in primary clarifiers:

EXAMPLE 1: Alum and polyelectrolyte addition at concentrations of 20 mg/l (as Al^{3+}) and 0.5 mg/l, respectively.

EXAMPLE 2: Ferric chloride and polyelectrolyte addition at concentrations of 20 mg/l (as Fe^{3+}) and 0.5 mg/l, respectively.

EXAMPLE 3: Lime precipitation using 150 mg/l of quick lime (CaO).

Capital costs were developed for a capacity of 1 mgd for the first two cases and 10 mgd for the lime precipitation method. The chemical treatment systems using alum and ferric chloride consist of bulk chemical storage facilities, transfer pumps, rapid mixing tanks, and flocculating basins. When lime is used, the treatment system includes a storage bin, lime feeder, and mixing tank. The costs are presented in Table 6-7.

Table 6-7

Capital Costs for Chemical-Addition Facilities
(ENR Index 1500)

Example	Capital Cost[1]
1 - Alum & Polyelectrolyte	$ 74,000
2 - Ferric Chloride & Polyelectrolyte	63,000
3 - Lime	150,000

[1] These costs contain no contingency for engineering design, bonding, and construction supervision.

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CHAPTER 7

EFFLUENT POLISHING TECHNIQUES

7.1 General

The use of effluent polishing for secondary effluent is a relatively new idea which is receiving increasing attention as a practical and economical method of upgrading to obtain increased organic and suspended solids removal from existing treatment facilities. It appears to be particularly applicable in those cases (and there are many) where it is necessary to increase efficiency by an overall amount of 10 to 20 percent in order to meet stricter water quality standards.

Four unit processes are considered in this manual for effluent polishing: 1) polishing lagoons; 2) microstraining; 3) filtration, including mixed, multi-media, and moving-bed filters; and 4) activated carbon adsorption. The reader is referred to the process design manual for suspended solids removal for an in-depth discussion of microstraining and filtration.

7.2 Polishing Lagoons

Polishing lagoons offer an opportunity for increased organic and solids removal at a minimum cost. There are two types of polishing lagoons which can be used, aerobic and facultative.

7.2.1 Aerobic Lagoons

Aerobic lagoons are generally subdivided into two groups:

1. Shallow lagoons, with depths in the range of 2.5 to 4.0 feet.

2. Deep lagoons, with aeration devices included to insure maintenance of aerobic conditions.

The shallow aerobic lagoon is one in which the algae-bacterial inter-relationship is optimized by providing as much light penetration as possible, and by maximizing photo-synthetic efficiency and bacterial oxidation of organic wastes. Operational data from a shallow aerobic lagoon are presented in Table 7-1 (1). The data indicate consistent BOD removals throughout the year, but marked increase in the concentration of suspended solids in the effluent during the summer periods, when algae activity is at its peak. The decreased solids removal without concurrent reduction of BOD removal during the summer months is caused by algae carryover in the effluent. This indicates that algae present in the effluent do not exert a significant amount of BOD demand during the five-day incubation used in the standard BOD test. The substantial increase in effluent suspended solids during the summer period, however, is a major disadvantage of the shallow lagoon as a dependable year-round polishing technique.

Table 7-1

Operational Data from a Shallow Aerobic Polishing Lagoon

Plant Location	Time Period	Flow	BOD lbs./day		BOD Removal	Surface Organic Loading	SS lbs./day		SS Removal	Remarks
			In	Out			In	Out		
		mgd			percent	lbs. BOD/acre day			percent	
Indian Creek, Kansas	1963 June 12-13	2.3	377	188	50.2	60.9	563	990	- 77	
	July 10-11	1.9	179	144	19.5	29.0	240	488	-103	Pond size - 6.2 acres
	Aug. 27-28	2.1	194	81	58.3	31.6	199	480	-140	Depth - 2.5 ft.
	Dec. 1-2	1.5	380	191	49.8	61.3	534	210	61	Odorous in Spring
	1964 Jan. 28-29	1.3	416	171	59.0	67.1	691	367	45	
	March 6-7	1.7	383	213	44.4	61.8	284	128	23	
	April 10-11	2.8	839	256	69.4	135	489	117	77	

Source: Loehr and Stephenson (1).

An alternative to the shallow lagoon is the deep, aerated, lagoon. These deeper lagoons can operate at greater surface organic loadings than shallow lagoons and yet maintain higher organic removals. Since oxygen is supplied to the basin by mechanical devices rather than furnished by the algae-bacterial biosymbiotic relationship, the algae production in the aerated lagoon is minimal compared to the shallow lagoon. Operational data for two aerated effluent polishing lagoons (8 to 10 feet deep) are presented in Table 7-2 (2).

Table 7-2

Removal Efficiencies for
Deep Aerated Effluent Polishing Lagoons

Plant Location	Surface Organic Loading lbs. BOD/acre/day	BOD Removal percent	SS Removal percent
Washington Borough, N.J.[1]	230	63	78
East Windsor Township, N.J.[2]	134	75	75

[1] Low-rate trickling filter plant
[2] Contact stabilization plant

Source: Hinde Engineering (2)

The lagoon at the Washington Borough Plant has average influent BOD and suspended solids concentrations of 43 mg/l and 70 mg/l, respectively. The average effluent BOD and suspended solids concentrations are 16 mg/l and 15 mg/l, respectively. The East Windsor Plant's polishing lagoon receives organic and solids concentrations as high as 80 mg/l, while the effluent concentrations are generally about 15 mg/l.

The deep aerated lagoon utilizes various types of aeration devices to supply the necessary oxygen to stabilize the organic matter. The aeration devices also must provide sufficient mixing to disperse oxygen uniformly and to prevent solids deposition. Because mechanical aeration devices supply considerably more oxygen per unit horsepower than air diffusion devices, a generalized approach for sizing mechanical aerators (floating or fixed) is discussed below.

Eckenfelder (3) has indicated that the power levels per 1,000 gallons of aeration tank capacity required to maintain solids under suspension and to disperse oxygen uniformly throughout the basin are 0.02 to 0.03 hp/1,000 gallons and 0.006 - 0.01 hp/1,000 gallons, respectively.

Edde (4) studied the degree of mixing provided by mechanical aerators used in treating wastewater from pulp mills. His study indicates that a velocity greater than 0.4 ft./sec.

should be maintained in the basin to prevent solids deposition, and that mixing energy input varies with the size of the aeration unit. The following values were given as sufficient mixing energy to disperse oxygen uniformly throughout the basin (4):

Table 7-3

Mechanical Mixing Energy Required for Oxygen Dispersion

Size of Aerators hp	Mixing Energy hp/1,000 gal.
100	0.014
50	0.018
20	0.021

The above discussions indicate that mechanical aerators can be designed to provide either complete mixing of solids including oxygen dispersion, or just to provide uniformly dispersed oxygen. In the latter case, solids deposition will occur in the basin.

Based on the reported organic surface loadings, the approximate land requirements for treating secondary effluent are as follows:

Table 7-4
Lagoon Land Requirements

Type	Land Requirements acres/mgd
Shallow aerobic lagoons	4.0
Deep aerated lagoons	1.0

7.2.2 Facultative Lagoons

Facultative lagoons are characterized by two distinct zones - aerobic and anaerobic. Hydraulic and organic loadings are such that the dissolved oxygen in the lower section of the lagoon is depleted but an aerobic layer is maintained near the surface. A cross-section of a typical facultative lagoon is shown in Figure 7-1.

At Peoria, Illinois, Fall (5) investigated the efficiency of a 10-foot deep polishing lagoon operated for 9-month periods each as a facultative lagoon and as an aerated lagoon. The results of his work are summarized in Table 7-5. It is interesting to note that both the BOD and the suspended solids concentrations in the effluent did not change appreciably during the period when the lagoon was operated aerobically as compared to the facultative operation. Fall also has stated that during the two winters of operation there was no

FIGURE 7-1
TYPICAL CROSS SECTION OF A FACULTATIVE LAGOON

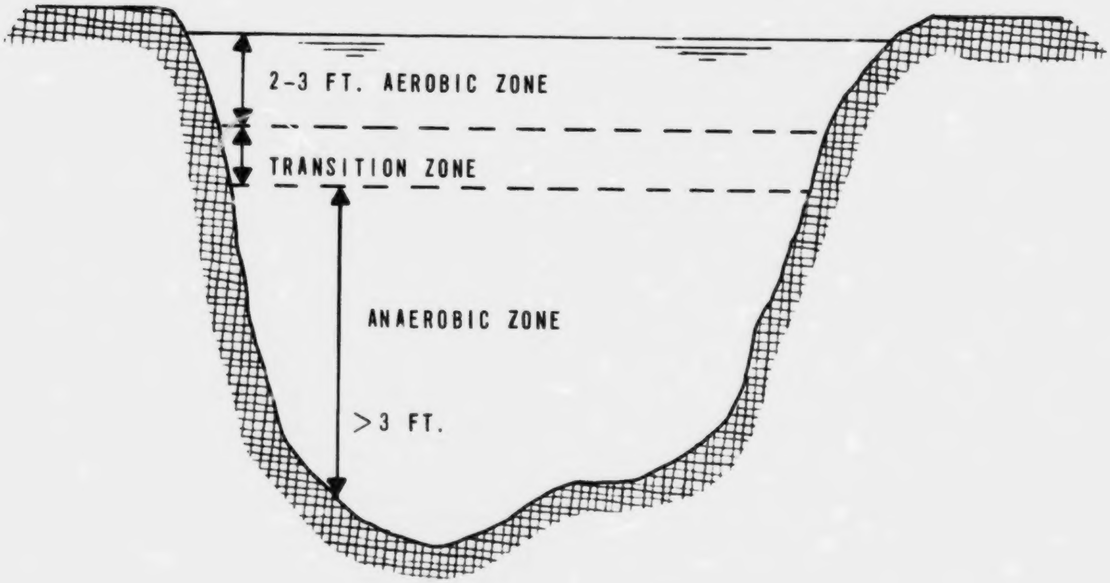

2-3 FT. AEROBIC ZONE
TRANSITION ZONE
ANAEROBIC ZONE
>3 FT.

Table 7-5

Comparison of Operational Data from Facultative and Aerated Polishing Lagoons

Description	Peoria, Illinois		Decatur, Illinois	Springfield, Missouri
	Aerated	Facultative	Facultative	Facultative
Type of Secondary Plant	Activated Sludge	Activated Sludge	Trickling Filter	Activated Sludge
Flow, mgd	0.66	0.65	6.8	18.7
Lagoon Size, acres	0.45	0.45	8.4	10
Average Pond Depth, feet	10	10	5.5	12
Influent BOD, mg/l	58	62	30	83
Effluent BOD, mg/l	34	30	18	30
Percent BOD Removal	41	52	40	64
Influent SS, mg/l	55	55	61	69
Effluent SS, mg/l	17	18	31	26
Percent SS Removal	67	67	49	62
Detention Time, days	1.8	1.82	2.3	1.61
Organic Surface Loading, lbs. BOD/acre/day	710	747	218	1,292
Air Applied, cu.ft. air/lb. BOD applied	223	0	0	0
Odor	None	None	None	—
Minimum Temperature of Lagoon During Study, °F	48	48	52	—

Sources: Peoria - Fall (5)
Decatur - Reynolds (6)
Springfield - Hickman (7)

ice on the pond. The lowest temperature of the pond effluent was 48°F, and this was recorded after 5 days during which temperatures were below 0°F. Facultative operation of the lagoon produced small amounts of algae in the pond during the summer period, but no odor problems were noted during the operation of this lagoon.

Operational data from the facultative effluent polishing lagoon in Decatur, Ill., also shown in Table 7-5, indicate that BOD and suspended solids removals averaged 40 percent and 49 percent, respectively, while operating under an organic surface loading of 218 lbs BOD/acre/day (6). As seen in Table 7-5, the facultative lagoon at Springfield, Mo., receives much higher surface organic loadings (approximately 1,290 lbs. BOD/acre/day) and still performs creditably, with average BOD and suspended solids removals of 64 and 62 percent, respectively (7).

A major disadvantage of using a facultative lagoon is the fact that the effluent will have a minimal dissolved oxygen content. Springfield, Mo., solved this problem by using cascade aeration (See Chapter 8, Section 8.2 on Post-Aeration). The effluent from the polishing lagoon flows over a series of 5 weirs with a total drop of 75 inches. The average dissolved oxygen in the effluent (September, 1970 through March, 1971) was 7.0 mg/l with a minimum and maximum, respectively, of 4.0 and 9.9 mg/l (7).

7.3 Microstraining

Microstraining has application in effluent polishing chiefly as a method of removing additional suspended solids (and their associated BOD) from wastewater treatment plant effluents. A typical unit is shown in Figure 7-2. (8). The microstrainer consists of a rotating drum with a peripheral screen. Influent wastewater enters the drum internally and passes radially outward through the screen, with deposition of solids on the inner surface of the drum screen. At the top of the drum pressure jets remove the deposited solids. This backwash water is then collected and returned to the head of the plant.

The screens employed in microstrainers have extremely small openings and are made from a variety of metals and plastics. Individual manufacturers have specific designs and sizes for the particular needs of any potential installation. One manufacturer offers the following grades of microfabric (9):

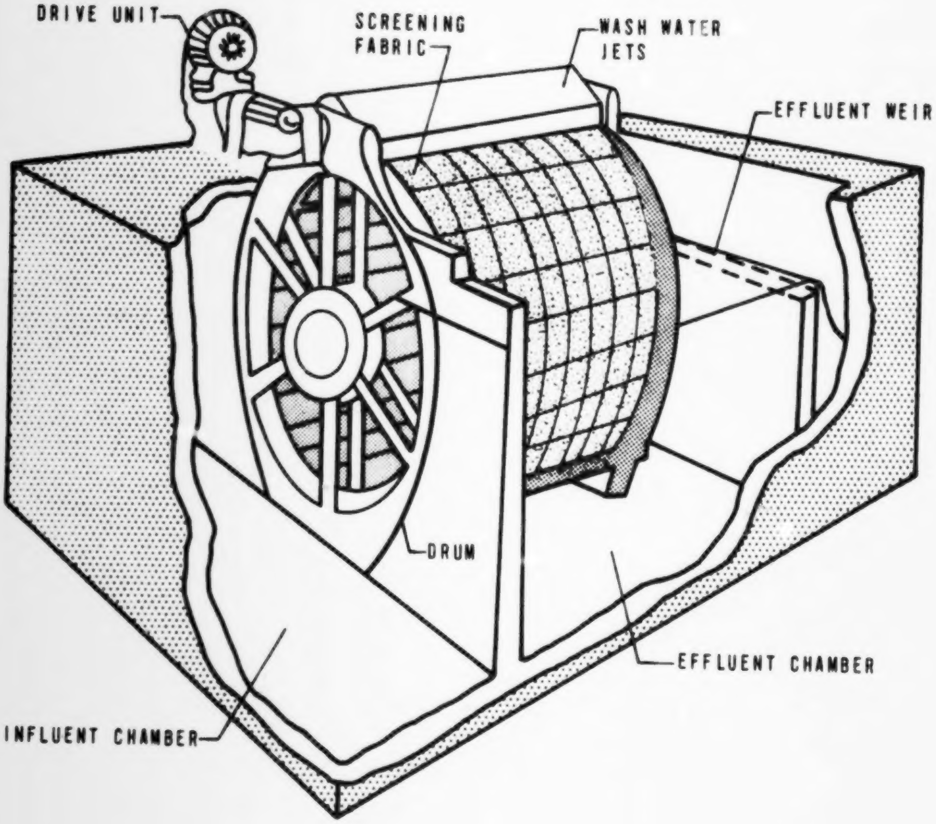
Table 7-6

Microstrainer Fabric Sizes

Opening microns	No. of Openings per sq.in.
23	165,000
35	80,000
60	60,000

FIGURE 7-2

TYPICAL MICROSTRAINER UNIT (8)


DRIVE UNIT
SCREENING FABRIC
WASH WATER JETS
EFFLUENT WEIR
DRUM
EFFLUENT CHAMBER
INFLUENT CHAMBER

The weave and shape of individual fabric wires are such that they allow the water from the backwashing jets to penetrate the screen and remove the solids mat which forms on the inside of the screen during its passage through the feed stream. Bodien and Stenburg (8) have noted that only about one-half of the applied washwater actually penetrates the screen; the rest flows down the outer perimeter into the effluent chamber. Previously-strained effluent can be used as washwater.

Although the microstrainers have small openings, the openings themselves cannot account for the removal efficiency of the unit. Actually, the mat of previously trapped solids provides the fine filtration which characterizes the unit performance. This being the case, Lynam et al (10) showed quantitatively that the slower the rate of drum rotation, the better the product water. Another factor which becomes important in light of this mat phenomenon is the nature of the solids applied to the microstraining process. For example Lynam et al were unsuccessful in filtering the resulting chemical floc when secondary effluent was coagulated ahead of the microstrainer unit.

As a section of the screen passes through its cycle, it becomes clogged rapidly as the solids mat forms. The continuous cleansing afforded by the backwashing jets at the apex of its travel must be augmented in some way to prevent the buildup of screen-clogging slimes over a period of time. Ultraviolet light placed in close proximity to the screen has been somewhat successful in slowing the development of these slimes. In general, however, units must be taken out of service on a regular basis (once/week, for example) to have the metal screens cleaned with a chlorine solution. In some instances, a similar cleaning with an acid solution may be required on occasion to deal with iron or manganese buildup on the metal screen. In cases where oil and grease problems occur, a hot water and/or steam treatment can be used to remove these materials from the screen.

One of the advantages of using a microstrainer is its low head requirement. It is, therefore, advantageous to transfer secondary effluent, without pumping, to a tertiary microstraining installation in order to minimize the shear forced imparted to the fragile biological floc. Head loss through the microstraining unit, including inlet and outlet structures is about 12 to 18 inches (9). Across the screen, a 6-inch limit is usually imposed at peak flows. Head losses in excess of this value are prevented by bypass weirs. Head loss buildup is reduced by increasing the rate of drum rotation and by increasing the pressure and flow of the backwashing jets. These adjustments can be made manually or automatically.

Other operating parameters include the hydraulic and solids loading on the unit. Lynam et al (10) found that the solids loading was the limiting factor in microstraining of activated sludge secondary effluent. Maximum capacity was found to be 0.88 lbs./day/sq.ft. at a hydraulic loading of 6.6 gpm/sq.ft. Excessively high solids during upset periods can reduce the capacity of the unit drastically from design levels. Chlorination immediately ahead of microscreening units should be avoided to protect the screens.

Operational data from various installations are presented in Table 7-7. The microstrainers using 23-micron fabric exhibited average solids removals ranging from 57 to 89 percent,

Table 7-7

Microstrainer Operational Data

Location	Plant Size mgd	Feed Type	Fabric Opening microns	SS Removal percent	Effluent SS mg/l	BOD Removal percent	Effluent BOD mg/l	Backwash % of flow	Reference
Brampton, Ontario	0.1	A.S.[1] Effluent	23	57	---	54	---	---	9
Lebanon, Ohio	Pilot	A.S. Effluent	23	89	1.9	81	---	5.3	8
	Pilot	A.S. Effluent	35	73	7.3	61	---	5.0	8
Chicago, Illinois	3.0	A.S. Effluent	23	71	≈3.0	74	≈3.0	3.0	10
Luton, England	3.6	Effluent from A.S. and T.F.[2]	35	55	7.3	30	---	3.0	11
Bracknell, England	7.2	T.F. Effluent	35	66	5.7	32	8.4	---	9,11

[1] A.S. - Activated Sludge

[2] T.F. - Trickling Filter

while the 35-micron fabric exhibited removals of 55 to 73 percent. In practice, the coarser 35-micron fabric is generally used for the removal of coarse solids. Maintenance of microstrainers can be quite costly, since most units will require cleaning at least once a week as previously mentioned.

Typical design parameters for microstrainers are presented in Table 7-8. These parameters must be evaluated to determine the proper microstrainer design for existing conditions.

Table 7-8

Typical Microstrainer Design Parameters

Parameter	Value
Drum Speed, rpm	0.7 - 4.3
Filter Fabric, microns	23 and 35
Average Hydraulic Loading[1], gal/sq.ft./hr.	
23-micron fabric	600
35-micron fabric	800
Backwash Pressure, psi	20 - 80
Amount of Backwash Water, percent of average flow	3 - 6
Maximum Hydraulic Loss through Screen, inches	6

[1] Based on submerged screen area.

Sources: Diaper (9) and Lynam (10).

7.4 Multi-media, Coarse-media, and Moving-bed Filters

Historically, sand filtration has not been an efficient method of polishing secondary treatment plant effluent because of low application rates, high head losses, and the need for frequent backwashing. This is largely because the normal backwashing of a sand filter results in a size-graded filter with the finest grains in the upper layers. The resulting stratification removes the bulk of the suspended matter in the upper levels, with a consequent inefficient use of the remaining depth of the filter.

However, developments in mixed, multi-media, and deep-bed coarse-media filters have necessitated a re-evaluation of the role of filtration in effluent polishing. In general, these modifications permit deeper penetration of the media by the suspended and colloidal contaminants; thus, there is a more effective utilization of the filter depth as compared to conventional sand filters. The increased utilization of filter depth is somewhat offset by the fact that increased backwashing rates and larger quantities of washwater are required to backwash the media properly.

Deep-bed coarse-media sand filters are a modification of the typical rapid sand filter. The deep-bed filter has a minimum of 4 feet of media as compared to the 2.5 feet or less of the usual rapid sand filter. The media in a deep-bed filter will generally range between 1 and 3 mm. in diameter, while the media in the rapid sand filter typically are less than 1 mm. in diameter.

In addition to coarse, mixed, and multi-media filters, a new filtering technique known as a moving-bed filter (MBF) has been developed by Johns-Manville Corporation (12). A schematic of the MBF is shown in Figure 7-3. The unit is basically a sand filter, but as the filter surface becomes clogged, the filtering medium is moved forward by means of a hydraulically-actuated mechanical diaphragm The clogged filter surface is removed mechanically or by gravity, to the extent that a fresh and clean filtering surface is exposed to the incoming chemically treated liquid. The unit is thus a form of countercurrent extraction device which has the capability of functioning on a continuous basis and does not have to be taken off stream for cleaning or backwashing.

The sand and accumulated sludge fall into a hopper and are washed and separated. The sand is then returned to the base of the filter unit.

Table 7-9 contains operational data from some of the various filtration processes previously discussed. Based on the available data, the mixed and coarse-media filters have significantly higher application rates while still maintaining a high degree of solids removal. However, it must be pointed out that the data for the MBF were developed for a phosphate removal study and, therefore, if BOD and suspended solids removals were the only considerations, the alum dosages could probably be reduced.

The use of pilot studies in the design of filtration units is recommended because of the numerous variables which govern the efficiency of the filtration processes. Some of the variables affecting filtration are: media depth, grain size, grain material, rate of filtration, in-flow solids concentration, characteristics of the suspension, water temperature, head loss, and backwash requirements.

FIGURE 7-3

SCHEMATIC OF THE MOVING-BED FILTER (MBF) (12)

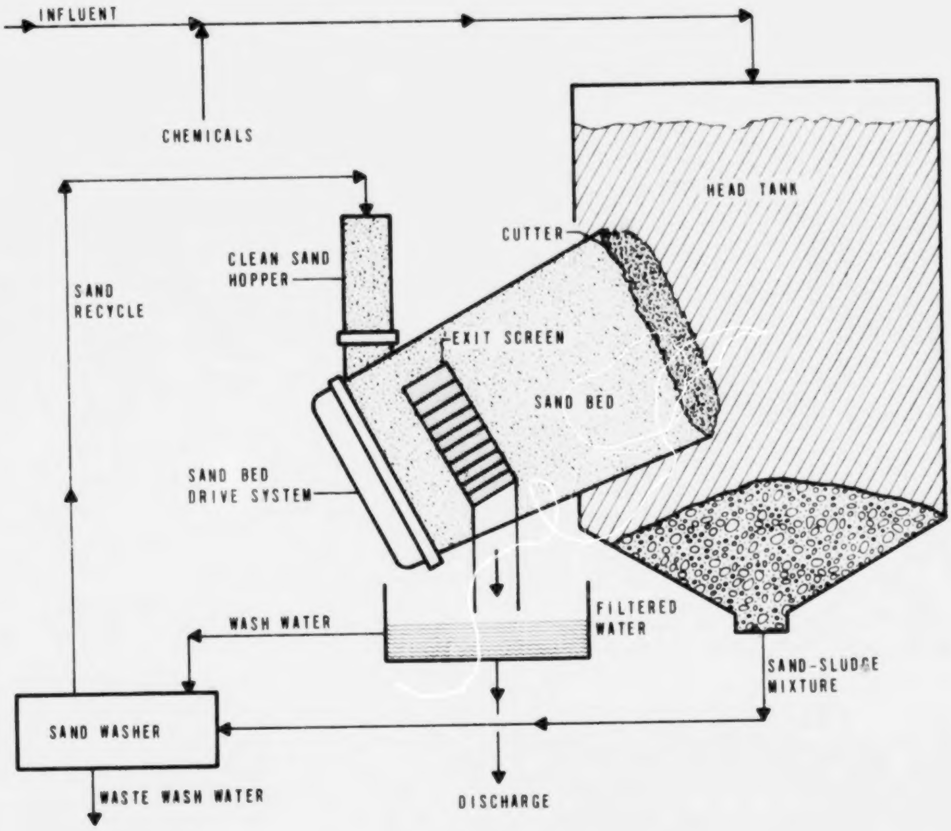

INFLUENT
CHEMICALS
HEAD TANK
CUTTER
CLEAN SAND HOPPER
SAND RECYCLE
EXIT SCREEN
SAND BED
SAND BED DRIVE SYSTEM
FILTERED WATER
WASH WATER
SAND-SLUDGE MIXTURE
SAND WASHER
WASTE WASH WATER
DISCHARGE

Table 7-9

Effluent Polishing Results - Filtration

Filter Type	Feed Type	Media Size mm.	Filter Depth ft.	Hydraulic Loading gpm/sq.ft.	SS Removal percent	Effluent SS mg/l	BOD Removal percent	Effluent BOD mg/l	Reference
Deep-Bed Coarse-Media									
Gravity downflow	T.F.[1] Effluent	1.0-2.0	-----	6	70	5.7	55	-----	13
Gravity downflow	T.F.[1] Effluent	0.9-1.7	2-3	3	67	-----	58	-----	14
Pressure upflow	T.F.[1] Effluent	0.9-1.7	5	3	85.5	5.0	74	2.5	14
Pressure upflow	A.S.[2] Effluent	0.9-1.7	5	3	77	-----	----	-----	14
Nepture Microfloc Mixed-Media	E.A.[3] Effluent	0.25-2	2.5	5	74	-----	88	-----	15
Moving-Bed	T.F.[4] Effluent	0.6-0.8	4.2	2	47	-----	71	-----	12
	T.F.[5] Effluent	0.6-0.8	4.2	2	67	-----	80	-----	12

[1] T.F. - Trickling Filter
[2] A.S. - Activated Sludge
[3] E.A. - Extended Aeration
[4] 100 mg/l alum and 0.2-0.75 mg/l anionic polymer
[5] 200 mg/l alum and 0.2-0.75 mg/l anionic polymer

7.5 Activated Carbon Adsorption

The limitations of conventional biological treatment processes in regard to reliable achievement of a high degree of organic removal (particularly of certain compounds which are refractory to biodegradation), along with increasingly strict water quality standards, emphasize the need for a supplementary organic removal process. Thus, activated carbon is presently being used to provide tertiary treatment of biologically treated effluents. Moreover, experience gained from the operation of activated carbon plants for tertiary treatment of wastewater suggests that activated carbon need not be restricted to a polishing role, but can be used as an alternative to biological treatment. Replacement of conventional biological treatment by activated carbon (i.e. secondary treatment application) is emphasized in the EPA process design manual for carbon adsorption. The discussion in the following pages is concerned exclusively with the tertiary application of carbon.

Activated carbon for wastewater treatment can be used either in the powdered or in the granular forms. The impracticality of economical regeneration has restricted the use of powdered carbon in wastewater treatment, although this problem is being resolved. Consequently, since the use of powdered activated carbon in wastewater treatment is not widespread, the discussion in the following section is limited to granular activated carbon.

7.5.1 Process Principles and Design Factors

The adsorption of organic materials from wastewater onto the activated carbon involves complex physical and chemical interactions. Biological degradation of adsorbed materials also occurs, and this can significantly enhance the overall treatment performance (16) (17).

The ability of activated carbon to adsorb large quantities of dissolved materials from wastewater is due to its highly porous structure and to the resulting large surface area, which provides many sites for adsorption of dissolved materials.

Important factors in the design of activated carbon treatment facilities include: pretreatment requirements; particle size; hydraulic loading and contact time; regeneration losses; flow configuration; and required effluent quality.

7.5.1.1 Pretreatment Requirements

Treatment of wastewater by activated carbon requires that the influent total suspended solids concentration be less than about 50 mg/l. This is essential in order to use the activated carbon bed as an adsorption medium and to minimize its filtration function. If the solids loading is much higher than 50 mg/l, a filter may be needed in advance of downflow carbon beds, or upflow carbon beds may be required for feasible operation.

7.5.1.2 Particle Size

Theoretically, carbon particle size primarily affects the rate of adsorption and not the capacity of the carbon. Adsorption rates are greater for smaller particle sizes than for larger particle sizes. However, adsorbents close to saturation will be less affected by particle size than adsorbents in their virgin state (16).

Data from Lake Tahoe indicate that there will be a reduction in adsorption capacity of about 20 to 35 percent in going from 12 x 40 mesh carbon to 8 x 30 mesh carbon at a relatively short contact time (16). This apparent difference in adsorption capacity attributable to particle size is minimized at longer contact times (18). Since finer particle sizes are susceptible to greater head losses, 12 x 40 mesh carbon is probably not suitable for use in downflow columns (18).

7.5.1.3 Hydraulic Loading Rate and Contact Time

Contact time, hydraulic loadings, and bed depth are interrelated physical parameters. Of the three, contact time is clearly the most important. Since the activated carbon treatment of wastewater requires that a definite contact time be established to complete the adsorption process, any increase in applied hydraulic load necessitates a deeper carbon column to maintain the same contact time.

Data obtained at the Pomona, California Pilot Plant indicate that Total Organic Carbon (TOC) removal does not vary significantly after fifteen minutes contact time for hydraulic loading rates of 4, 7, and 10 gal./min./sq.ft. (16). It was further noted that for equivalent contact times, the percent TOC removal was similar for hydraulic loading rates of 4, 7 and 10 gal./min./sq.ft. These results indicate that contact time is more important than applied hydraulic loadings, and is in fact the most important design factor in carbon adsorption systems.

Typical hydraulic loading rates and contact times used in various locations are shown in the following table:

Table 7-10

Typical Hydraulic Loading Rates and Contact Times

Plant Location	Hydraulic Loading Rate gpm/sq.ft.	Contact Time[1] minutes	Type of Treatment
Pomona	7	36	Tertiary
Lake Tahoe	8	18	Tertiary
Nassau County	7.5	24	Tertiary

[1] Empty Bed

Source: Zanitch and Morand (19)

It should be noted that both gravity and pumped flow systems are available; gravity flow systems are not likely to be practical at hydraulic loading rates above about 4 gpm/sq.ft.

7.5.1.4 Effect of Regeneration

Activated carbon requires regeneration when its adsorption capacity is exhausted. Considerable effort has been expended to determine the effect of regeneration on adsorption capacity of the carbon. However, since few research groups have regeneration facilities, only limited data are available. Results obtained at Pomona (16) indicate that the adsorptive capacity decreases by approximately 35 percent after 7 regeneration cycles, as indicated in Figure 7-4. It was also determined that regeneration does not affect the degree of organic removal in subsequent exhaustion cycles. This loss of capacity is not necessarily a critical factor, since it is necessary to make up physical losses of carbon after each regeneration cycle. These losses are caused by several factors: carbon is burned and lost through the stack as combustion products; or is abraded into dust in the course of handling. Further "losses" are due to the buildup of inorganic ash in the carbon particles during repeated use and regeneration.

7.5.1.5 Flow Configuration

Depending on the dissolved organic and suspended solids loading, any of several optional flow configurations can be adopted:

1. Downflow Beds in Series - the lead contactor is removed, regenerated, and replaced in line at the downstream end, the other contractors being moved up in sequence.

2. Downflow Beds in Parallel - parallel beds are arranged in a staggered exhaustion pattern so that when one is exhausted and removed from service, the product of the others can be blended with that portion of flow normally treated by the exhausted contactor to maintain the required product quality for the entire plant.

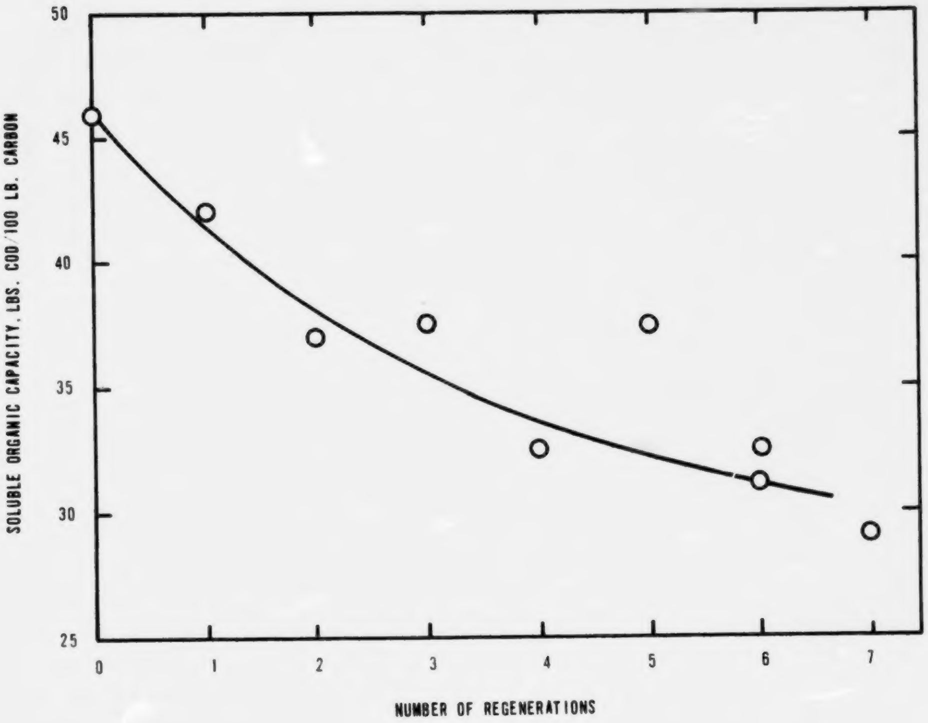
3. Upflow Beds (expanded or partially expanded) - no head loss is built up, and no backwashing is necessary; post-filtration is required; the same series and parallel considerations apply as for downflow operation.

4. Upflow (moving bed) - exhausted lower strata of the bed are continuously removed and replaced (at the top of the bed) by virgin carbon.

Downflow beds always require backwashing unless a pre-filtration step is added. Upflow beds do not require backwashing since no head losses build up; however, post-filtration is necessary.

FIGURE 7-4

EFFECT OF REACTIVATION ON ADSORPTION CAPACITY (17)



7.5.1.6 Effluent Quality

In addition to the above design considerations, the question of effluent quality standards should not be neglected. It should be clear at this point that the carbon adsorption process can be readily controlled and designed to achieve virtually any desired organic removal efficiency. It is probably unique among tertiary processes in this respect.

7.5.2 Laboratory and/or Pilot Plant Investigations

Activated carbon removes dissolved materials from wastewaters by a combination of three mechanisms: adsorption, filtration, and biological degradation. Therefore, in order to judge the effectiveness of activated carbon for wastewater treatment, both laboratory and pilot testing are required.

The adsorption mechanism can be evaluated in the laboratory by running adsorption isotherms. Actual plant conditions should be simulated with regard to temperature, pH, and pre-treatment. A detailed isotherm procedure is given in many books, as well as in the above mentioned carbon design manual. Only a brief description will be given here.

Adsorption isotherms are normally conducted by contacting a sample of wastewater with varying amounts of pulverized carbon for a standard interval of time. The wastewater sample is analyzed for TOC, COD, or BOD (as deemed necessary), both before and after contacting with the pulverized activated carbon. The treated water should be coarse filtered prior to analysis to eliminate carbon fines. The isotherm is a plot of the amount of solute adsorbed per unit weight of carbon as a function of residual concentration of solute. The isotherm is empirically represented by the following expression (14):

$$x/m = KC^{1/n}$$

where:

x = weight of solute adsorbed
m = weight of carbon
C = equilibrium concentration of solute in solution after adsorption
K and n are constants

The isotherms are normally plotted on a log-log scale. The extrapolation of the isotherm line to the initial concentration (abscissa) gives the theoretical adsorption capacity of that carbon when it is in equilibrium with the influent concentration.

The advantage of isotherms are; 1) they are relatively simple tests to perform; 2) they indicate whether the desired degree of treatment can be readily achieved; and 3) they give the approximate adsorptive capacity of the carbon in a column application. However, isotherm results should not be used to extrapolate carbon capacities and dosages to full-scale plant size.

Typical isotherms obtained for the same carbon with different secondary wastewater effluents were reported by Masse (20) (see Figure 7-5). The results shown in Figure 7-5 indicate that the adsorptive capacity of carbon with respect to COD varies from 0.37 to 0.42 lb. of COD per lb. of carbon. This is equivalent to 0.22 to 0.25 lb. of carbon/1,000 gallons of throughput. The carbon requirements (per unit volume of wastewater treated) obtained from isotherms are conservative (i.e. high), because removal by adsorption alone is estimated.

Since isotherms cannot measure the quantity of organics removed by filtration and biological action, pilot column testing must be conducted to evaluate the effect of these factors. Column testing helps to determine: 1) the required contact time; 2) the adsorptive capacity of the carbon; 3) the pressure drop across the beds and backwash requirements for downflow operation; and 4) the shape of the column exhaustion wave front.

The column used in pilot testing should have a diameter of at least 4 inches. The depth of column depends on the range of contact times being considered, as does the hydraulic loading. Normally, 2 to 4 columns are used in series, since this arrangement permits evaluation of the effect of different contact times on effluent quality. When loading granular carbon into the test column (a "wet" packing procedure is recommended), care must be exercised to avoid entrapping air within the carbon column. Air entrapment causes channeling and reduces the contact area, which in turn yields false test results. The process design manual for carbon adsorption describes the conduct of pilot operations in more detail.

7.5.3 Uses of Activated Carbon for Upgrading Treatment Plants

Activated carbon has been used successfully to provide tertiary treatment at three locations: Lake Tahoe, Pomona, and Nassau County. Experience at these places clearly demonstrates the ability of activated carbon to produce effluents with very low levels of organics. At Lake Tahoe, the secondary effluent is treated with lime followed by clarification and mixed-media filtration prior to treatment with the activated carbon; at Pomona, secondary effluent is treated directly in activated carbon columns; in Nassau County, secondary effluent is alum-clarified prior to treatment in activated carbon columns. Some operating results as well as design parameters for these carbon column installations are shown in Table 7-11.

7.6 Process Designs and Cost Estimates

A cost comparison has been prepared for various effluent polishing processes, and the results are presented in Table 7-12. The cost information on the MBF was obtained from a phosphate removal study (12) and, therefore, is higher than would be expected for an application not including phosphorus removal. The data on microstraining and sand filters were taken from Smith and McMichael (22) and updated to an ENR index of 1500. Process designs and cost estimates for tertiary treatment by granular activated carbon have been examined in an earlier EPA publication, prepared by the M.A. Kellog Co. and Swindell-Dressler Co., Divisions of Pullman, Inc., (16).

FIGURE 7-5

COD ISOTHERMS USING VIRGIN CARBON
AND DIFFERENT SECONDARY WASTEWATER EFFLUENTS (20)

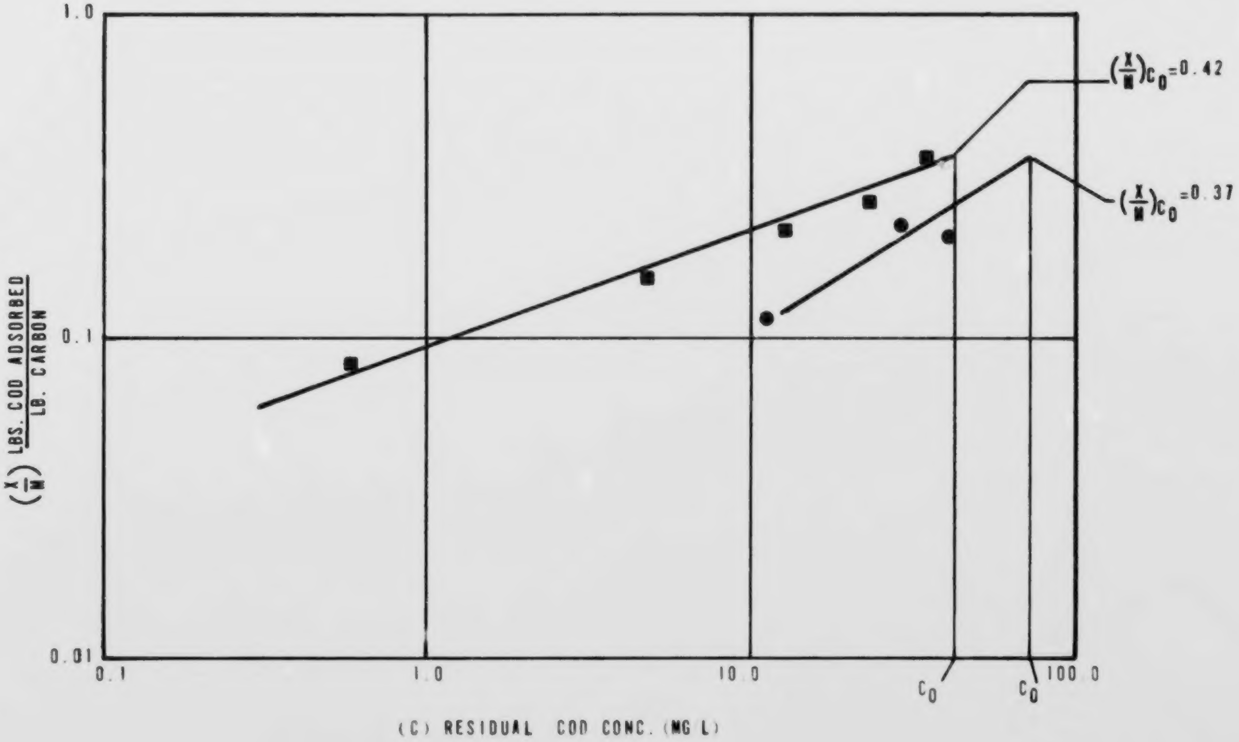


Table 7-11

Summary of Operating Results Using Activated Carbon for Tertiary Treatment

	Pomona		Lake Tahoe		Nassau County	
Operating Data						
Capacity	200 gpm		1,800 gpm		400 gpm	
Source of Waste	Domestic		Domestic		Domestic	
Secondary Treatment	Standard Activated Sludge		Standard Activated Sludge		High-rate Activated Sludge	
Pre-treatment	Chlorination		Coagulation and Filtration		Coagulation and Filtration	
Carbon Type	16 x 40 mesh		8 x 30 mesh		8 x 30 mesh	
Column Configuration	4-Stage Downflow		2-Upflow in Parallel		4-Stage Downflow	
Column Dimensions	6' dia. x 9'deep		12' dia. x 14' deep		8' dia. x 6' deep	
Nominal Contact Time	36 minutes		13 minutes		24 minutes	
Loading Rate	7 gpm/sq.ft.		8 gpm/sq.ft.		7.5 gpm/sq.ft.	
Carbon Column Performance	Influent	Effluent	Influent	Effluent	Influent	Effluent
COD, mg/l	47	10	20-30	2-10	–	5
BOD, mg/l	–	--	5-20	2-5	--	–
Color, Pt-Co Units	30	3	20-50	5	–	–
Carbon Dosage	350 lbs/million gallon		250 lbs/million gallon		500 lbs/million gallon	

Source: Zanitch and Morand (19)

Table 7-12

Economic Comparison of Various Effluent Polishing Processes
(ENR Index 1500)

	Capital Costs Thousands of Dollars					Yearly Costs[1] cents/1,000 gallons			
Design Flow mgd	Deep-[2] Aerated Lagoon	Micro-[3] straining	Sand[3] Filters	MBF[4]	Multi-Media[5] Filters	Deep[6] Aerated Lagoon	Micro-[3] straining	Sand[3] Filters	MBF[4]
1.0	—	39	120	302	80	—	4.5	7.6	14
4.0	270	135	290	—	200	2.0	3.9	4.6	—
10.0	—	300	530	—	350	—	3.7	3.4	—

[1] Yearly costs include: amortization and interest at 4.5 percent for 25 years, operating and maintenance costs.
[2] Capital cost data obtained from Pierce (21).
[3] Cost data obtained from Smith and McMichael (22).
[4] MBF cost data obtained from Bell, et al (12) include operation and maintenance costs.
[5] Capital cost data obtained from Neptune Microfloc (23).
[6] Costs estimated by using information furnished in Pierce (21).

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CHAPTER 8

PRE-AERATION AND POST-AERATION PRACTICES

8.1 Pre-Aeration

Pre-aeration of wastewater has been practiced for over 50 years throughout the United States, mainly for the purpose of odor control and/or the prevention of septicity. Initially, short aeration periods ranging up to 15 minutes were used. As aeration periods were lenghtened, the additional benefits of grease separation and flocculation of solids became evident (1).

Efforts have been made by Roe (1), and by Seidel and Baumann (2) to study the effects of pre-aeration on primary clarifier performance. In 1951, Roe tabulated operational data from 38 plants using pre-aeration. These results are presented in Figure 8-1 (1). The data in Figure 8-1 illustrate the effect of pre-aeration time on subsequent suspended solids (SS) removal in the primary clarifier. The quantities of air used varied between 0.06 and 0.15 cu.ft./gallon. Roe further found that in order to maintain proper agitation, plant operators varied the air supply between 1.0 and 4.0 cu.ft./lineal foot of tank, depending on the physical tank layout and type of aeration equipment used. BOD removal could not be correlated with the observed solids removal.

In 1961, Seidel and Baumann conducted a comparative study at the Ames, Iowa, secondary treatment plant to determine the effect of pre-aeration on primary clarifier performance. They determined that with 45 minutes detention and an aeration rate of 0.1 cu.ft. air/gallon, BOD and SS removals were increased by 7 to 8 percent in the primary tank. The 7 to 8 percent increased removals in the primary tank may or may not be realized on overall plant performance, e.g., part of the incremental solids removed as a result of pre-aeration might have been removed in the secondary clarifier without using pre-aeration.

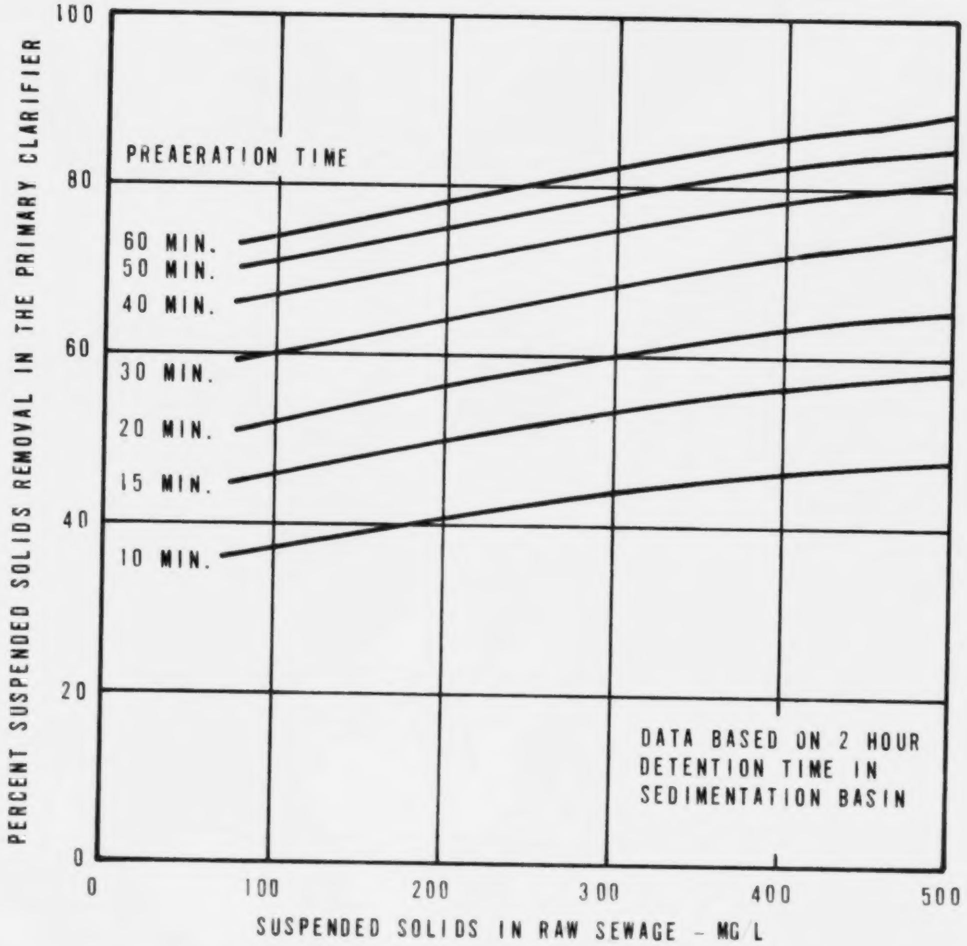
Seidel and Baumann also evaluated pre-aeration economics based on conventional design standards and found that the cost of pre-aeration increases the total annual operating cost by 2 to 3 percent. However, they felt a slight improvement in clarifier efficiency would narrow or eliminate the cost differential.

The Ten-States Standards recommend detention times of 30 minutes for effective solids flocculation and at least 45 minutes for appreciable BOD reduction with pre-aeration (3). In addition, it is stated that the use of polyelectrolytes may substantially reduce these detention times.

Although the use of aerated grit chambers is becoming increasingly popular as a pretreatment unit in wastewater treatment plants, their use should not be expected to substantially increase the BOD or SS removal in the primary clarifier.

FIGURE 8-1

EFFECT OF PREAERATION TIME ON SOLIDS REMOVAL (1)



8.1.1 Applicability to Plant Upgrading

Due to the limited amount of SS and BOD removal achieved as a result of pre-aeration, its use for the upgrading of solids and organic removal is limited. However, the use of pre-aeration for preventing septicity of raw wastewater and grease removal should be considered in the overall evaluation of pre-aeration as a unit process.

8.1.2 Process Design and Cost Estimates

Capital cost estimates for pre-aeration facilities for three plants with capacities of 1, 3, and 5 mgd are shown in Table 8-1:

Table 8-1

Capital Costs for Pre-Aeration Facilities[1]
(ENR 1500)

Plant Size mgd	Capital Costs for Pre-Aeration Facilities In Thousand Dollars
1	$ 75
3	148
5	215

---

[1] Does not include land costs, contingencies, engineering design or bonding.

The pre-aeration facilities were based on a detention time of 45 minutes and an air supply of 0.1 cu.ft./gallon. The basin was of reinforced concrete construction and the air requirement was supplied by floating mechanical aerators.

8.2 Post-Aeration

Many states are considering or have already enacted legislation requiring the maintenance of minimum dissolved oxygen concentrations in wastewater treatment plant effluents. This is required since most water quality standards specify a minimum dissolved oxygen (D.O.) concentration of 4.0 mg/l, while most secondary plants usually discharge effluents ranging between 0.5 and 2.0 mg/l.

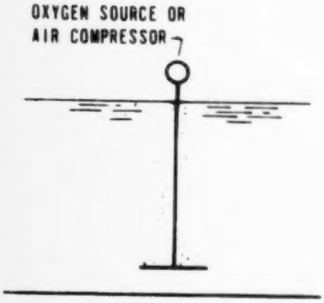
There are at least four methods available for the post-aeration of a wastewater treatment plant's effluent. These are shown in Figure 8-2. Most of these devices were initially developed for water treatment and are now being used in the wastewater treatment field.

8.2.1 Diffused Aeration

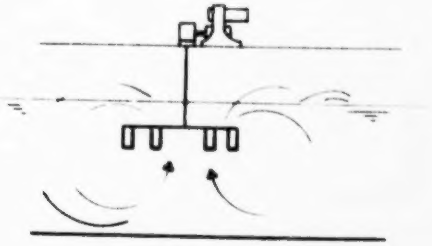
Diffusion aerators are usually placed in concrete tanks which are commonly 9 to 15 feet deep and 10 to 30 feet wide. Ratios of width to depth should not exceed two, if effective

FIGURE 8-2

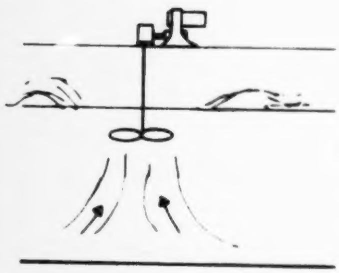
VARIOUS POST AERATION DEVICES


OXYGEN SOURCE OR
AIR COMPRESSOR

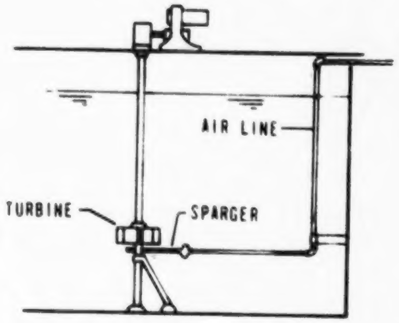
A. DIFFUSED AERATION



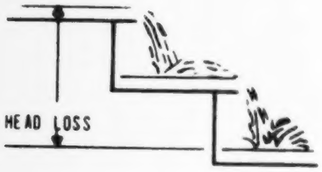
B-1 TURBINE TYPE AERATOR



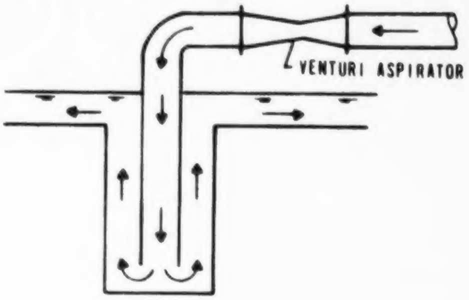
B-2 PUMP TYPE AERATOR


AIR LINE
TURBINE
SPARGER

B-3 AGITATOR SPARGED SYSTEM



C CASCADE AERATOR


VENTURI ASPIRATOR

D U TUBE AERATOR

mixing is to be obtained. Tank length is governed by the desired detention period, which usually varies from 10 to 30 minutes. The maximum air required is typically 0.15 cu.ft./gallon.

The use of oxygen aeration in the activated sludge process may eliminate the need for post-aeration. Oxygen-aerated mixed liquor discharged to the secondary clarifier usually has a D.O. of at least 6.0 mg/l (4).

8.2.2 Mechanical Aeration

Mechanical aerators are generally grouped in two broad categories: turbine types and pump types, as shown in Figure 8-2. In all types, oxygen transfer occurs through a vortexing action and/or from the interfacial exposure of large volumes of liquid sprayed over the surface.

To avoid interference between units, aerator manufacturers recommend a minimum basin size of 15 to 50 feet square and a minimum depth of 5 to 8 feet, depending on the horsepower of the aerator.

One aerator design equation proposed by Kormanik for a post-aeration basin is (5):

$$P = \frac{0.347\, Q\, (C - C_o + R_r t)}{N_o \left(\frac{C_{sw} - C}{C_{20}} \right) \theta^{T-20}\, a\, F_g\, \eta}$$

where:

P = Horsepower required
Q = Wastewater effluent flow, mgd
T = Design temperature of effluent, ^{o}C
N_o = O_2 transfer efficiency under standard conditions in tap water, lb. O_2/hp-hr.
F_g = Correction factor related to a change in basin geometry
C = Required final D.O. level after post-aeration, mg/l
C_o = D.O. concentration of the incoming wastewater effluent, mg/l
C_{sw} = O_2 saturation concentration of effluent, mg/l, where $C_{sw} = C_s \times \beta$
C_{20} = D.O. saturation of tap water at $20^{o}C$, mg/l
C_s = D.O. saturation of tap water at temperature T, mg/l
R_r = O_2 utilization rate, mg/l/min.
a = O_2 transfer coefficient of the effluent (alpha factor)
t = Detention time in minutes
η = Aerator efficiency correction
β = Oxygen saturation coefficient in wastewater (beta factor)
θ = Temperature coefficient varying between 1.02 and 1.024

8.2.3 Cascade Aeration

Cascade aeration takes advantage of the effluent discharge to create a series of steps or weirs over which the flow moves. The objective is the maximization of turbulence to increase oxygen transfer. Head requirements vary from 3 to 10 ft., depending upon the initial D.O. and the desired increase. If the necessary head is not available, effluent pumping is required.

In England, the Water Research Laboratory has performed investigations to qualify as much as possible the layout of cascade aeration schemes. Barrett and others proposed the following formulae (6):

$$r = C_s - C_a / C_s - C_b$$

$$r = 1 + 0.11 \; ab \; (1 + 0.046T) \;\; h$$

where:

r = The deficit ratio
C_s = Oxygen saturation value corresponding to temperature T, mg/l
C_a = Oxygen concentration above the weir, mg/l
C_b = Oxygen concentration below the weir, mg/l
a = Water quality parameter equal to 0.8 for a wastewater treatment plant effluent
b = Weir geometry parameter equal to unity for a free weir and 1.3 for the step weirs used in their experimental work
T = Water temperature in ^{o}C
h = Height in feet through which water falls

For example, to raise the D.O. concentration of a wastewater treatment plant effluent from 0.5 mg/l to 4.5 mg/l at 20^{o}C, the overall height requirement for a series of step weirs would be approximately 4.0 feet. However, it should be pointed out that the values of the parameters a and b are somewhat arbitrary and need further refinement to substantiate preliminary results.

8.2.4 U-Tube Aeration

The U-tube aerator consists of two basic components: a conduit to provide a vertical U-shaped flow path and a device for entraining air into the stream flow in the down leg of the conduit as indicated in Figure 8-2. The entrainment device is one of two types: 1) aspirator; or 2) compressor and diffuser. In either case, the entrained air is carried along the down leg of the tube because the water velocity exceeds the buoyant rising velocity of the air bubbles.

Various design considerations include air-to-water ratio, tube cross-sectional area, and depth. The maximum air-to-water ratio practicable is a function of the velocity through the system. At velocities of approximately 4 fps, 20 percent air-to-water injection is about the limit for satisfactory operation (7). The hydraulic head requirements for plants of 5 mgd or less should be less than 5 feet. If sufficient head is not available, the flow may be pumped through the U-tube.

Speece and Orosco (7) have suggested that one economic method of construction for deep U-tubes, greater than 20 feet in depth, would be a circular hole bored into the soil. The hole would be cased and a smaller pipe then suspended a few feet from the bottom of the hole as shown in Figure 8-2. The diameter of the smaller pipe is selected so that its cross-sectional area is approximately equal to the cross-sectional area of the annular space between the two pipes. Thus, the velocity of the water will be approximately equal in both legs of the U-tube.

Presently there are no known U-tube installations in wastewater treatment plants for post-aeration. However, their applicability is presently being investigated by EPA as a possible pre-aeration device in sewer lines. The possibilities of using a U-tube as a post-aeration device seem good at this time. The additional benefits of no moving parts and little or no associated labor and maintenance make the U-tube device extremely attractive.

Of all the types of post-aeration methods, it is likely that mechanical aeration and U-tube aeration will find extensive application in the future.

8.2.5 Process Designs and Cost Estimates

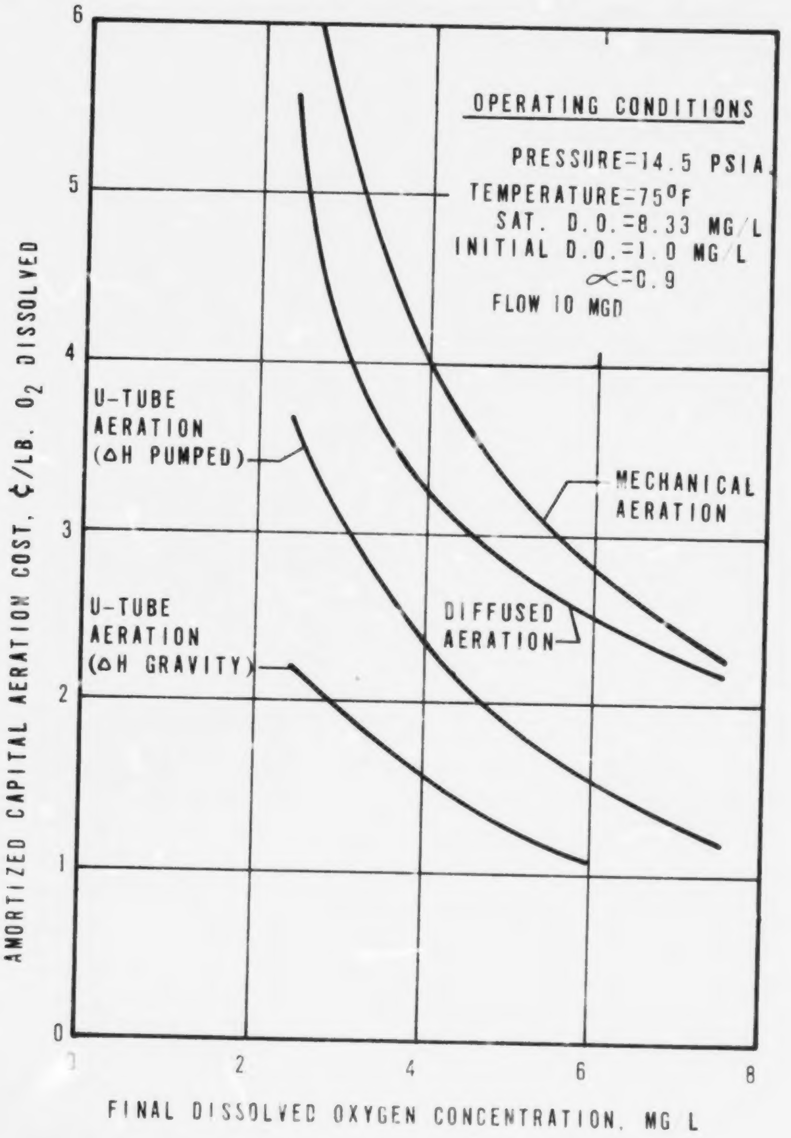
Mitchell and Lev (8) have prepared a cost comparison between mechanical, diffused, and U-tube aeration for post-aeration of a 10-mgd treatment plant effluent. This cost comparison is presented in Figure 8-3. The costs are expressed as amortized capital cost per pound of oxygen dissolved. The amortization was based on 4.5 percent for 25 years.

As shown in Figure 8-3, the gravity U-tube aeration device is the most economical for the particular conditions investigated. However, it must be cautioned that these amortized capital costs represent only 20 to 50 percent of the total yearly cost of operating these devices. Maintenance, operation, and power charges may be substantial. These costs were not included since there is a scarcity of reliable operating costs for aeration systems. However, when U-tube aeration can be operated under conditions where head is available, it is likely to be the cheapest of all devices currently on the market. This is due to the fact that U-tube aeration devices have low maintenance and power requirements.

FIGURE 8-3

CAPITAL COST COMPARISONS FOR POST AERATION OF SECONDARY TREATED EFFLUENT (8)

ENR=1500



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CHAPTER 9

DISINFECTION AND ODOR CONTROL

9.1 General

Disinfection and odor control are two areas which are receiving increased attention from regulatory agencies through the establishment and enforcement of rigid bacteriological effluent standards and air pollution standards.

9.2 Disinfection

In 1968, 41 percent of all municipal wastewater plants in the United States were using chlorination for disinfection purposes (1). Other disinfectants, ozone, for example, are currently being studied by EPA to evaluate their potential use in disinfecting wastewater treatment plant effluents.

The following table was taken from the Water Pollution Control Federation's Sewage Treatment Plant Design Manual and contains ranges of chlorine dosages recommended for disinfection (2):

Table 9-1

Chlorine Dosage Ranges

Waste	Chlorine Dosage mg/l
Raw Sewage	6 to 12
Raw Sewage (septic)	12 to 25
Settled Sewage	5 to 10
Settled Sewage (septic)	12 to 40
Chemical Precipitation Effluent	3 to 10
Trickling-Filter Effluent	3 to 10
Activated Sludge Effluent	2 to 8
Sand Filter Effluent	1 to 5

To be effective for disinfection purposes, a chlorine residual of 0.2 to 1.0 mg/l is recommended, with a contact time of not less than 15 minutes at peak flow rates (2).

It is possible to maximize the efficiency of an existing chlorine contact tank by improving the flow pattern through the tank. This could be accomplished by using baffles and by locating the chlorine addition point where complete mixing with the wastewater is assured. Beyond these modifications, it may not be possible to substantially increase the performance of chlorine contact tanks without increasing tank capacity to provide adequate detention time.

At the present time, ozone is being used for disinfection purposes in the water treatment field. Preliminary information indicates that ozone disinfection of potable water requires only 5 minutes of contact time to accomplish the same degree of disinfection as 15-minute contact with chlorine (3). In the future, it may be that ozone could be used effectively for wastewater disinfection purposes.

9.3 Odor Control

Wastewater treatment plants serving large municipalities are generally characterized by extensive collection systems with correspondingly high detention times. For example, the Washington, D.C. Pollution Control Plant serves areas as far as 25 to 30 miles away. This type of situation often leads to odor problems during summer periods. Odor problems are characteristically most critical during the plant's low flow periods (approximately 9 p.m. to 4 a.m.), due to increases in the sewer detention time.

9.3.1 Odor Generation and Characteristics

Odors from wastewater treatment plants can usually be attributed to three sources: septic raw wastewater, overloaded secondary treatment facilities, and sludge treatment practices.

Septicity in wastewaters is caused by the depletion of dissolved oxygen due to long residence in sewers and the subsequent increase in anaerobic activity. As wastewater becomes anaerobic, facultative and anaerobic bacteria flourish. These bacteria utilize nitrates and sulfates present in wastewater as their oxygen source. The reduction of sulfate ions produces the highly odorous gas, hydrogen sulfide. Other odorous gases which may be present are indole, skatole, mercaptans, disulfides, volatile fatty acids, and ammonia.

Increased summer temperature and extended sewer detention times can result in the rapid build-up of hydrogen sulfide and carbon dioxide according to the following reactions (4):

$$SO_4^{=} + \text{organic} \xrightarrow[\text{bacteria}]{\text{anaerobic}} S^{=} + H_2O + CO_2$$

$$S^{=} + 2H^{+} \rightleftharpoons H_2S$$

At a pH level below 8, the equilibrium shifts toward the formation of non-ionized H_2S and is about 80 percent complete at pH 7. At pH 8 and above, most of the reduced sulfur exists in solution as HS^- and $S^=$ ions (4). H_2S is noticeable even in the cold when present in water to the extent of 0.5 mg/l. When present to the extent of 1.0 mg/l, it becomes very offensive (5).

Overloaded secondary treatment facilities are also a potential source of odor. If the air supply to an activated sludge aeration tank is inadequate, odorous conditions usually develop. It is also possible that a properly sized air supply system can strip odorous gases from septic wastewater.

Odors associated with sludge treatment occur in thickening, digestion, and sludge dewatering facilities. Thickeners may receive both septic primary and secondary sludges. Gases from well-operated digesters may contain small quantities of H_2S, which are usually destroyed by normal flaring of digester gas. The predominant odor in digested sludge is ammonia, although traces of volatile organic acids may be present.

9.3.2 Odor Measurement

Odor data are generally qualitative rather than quantitative in nature. The two available quantitative methods are the H_2S determination and the Threshold Odor Number (TON). The latter method is only semi-quantitative in that determination of the TON is dependent on the olefactory senses of the individual performing the analysis. This can be de-personalized somewhat by using a panel to determine the TON value for a sample.

9.3.3 Odor Control Methods Available

The various methods available for control of odors emanating from a wastewater treatment plant are:

1. Changes in the operational procedures and new techniques.
2. Chemical treatment or pre-treatment, which might include chlorine, ozone, lime, or powdered carbon.
3. Collection and treatment of noxious gases.

9.3.3.1 Changes in Operational Procedures and New Techniques

Odors associated with septic wastewater are generally not amenable to solution through operational changes within the treatment plant itself. The applicability of in-sewer aeration methods for reduction of odors and hydrogen sulfide corrosion is currently being investigated by EPA. Among the procedures being evaluated are U-tube installations and pure oxygen injection into force mains.

Many sludge odors in a plant are a direct result of an improperly operated or overloaded anaerobic sludge digester. Improved temperature control and better mixing of digester contents may alleviate the odor problem

9.3.3.2 Chemical Treatment

Chlorination is probably the most widely used of the chemical treatment processes available to control odors because it is effective and most treatment operators have had experience in handling chlorine. The presence of chlorination facilities at the plant for disinfection is a further reason for its utilization.

Chlorination is used for two purposes: to retard biological action which produces odors, and to react chemically with odorous sulfur compounds, oxidizing them to innocuous sulfur forms, usually free colloidal sulfur.

Table 9-2 contains a summary of odor reduction data for a chlorinated raw domestic wastewater. A pre-chlorination dosage of 10 mg/l at maximum flow was recommended for odor control. This dosage is substantiated by the data in Table 9-1 (6). In addition, another incremental 5 mg/l of chlorine capacity was recommended as an adequate margin of safety for peaks in sulfide levels or chlorine demand.

Table 9-2

Effect of Chlorine on Odor Reduction For A Raw Domestic Wastewater (6)

Chlorine* Dosage mg/l	Detention Time in Minutes	
	1 Sulfide mg/l	120 Sulfide mg/l
0	1.7	1.7
5	0.5	0.7
10	0.2	0.2
25	0.2	0.2
50	0.2	0.2

*pH = 7
Temperature = 85°F

Ozone has been added to wastewaters for odor control with some favorable results. Because of the extremely high reactivity of ozone, a much higher ozone demand is generally exhibited by a wastewater than would be exhibited for chlorine. However, the use of oxygen aeration in secondary treatment may have an added benefit, since the exhaust gas could provide the ozone generator with an economic source of oxygen. Due to the high cost of ozone generation, the use of ozone for odor control may be limited (6).

Lime and powdered carbon have also been used in various applications for odor control. The addition of lime to septic wastewater raises the pH. Since the solubility of H_2S increases with increasing pH, less H_2S evolves, thereby decreasing the odor level. Powdered activated carbon adsorbs odor-causing materials and, thereby, decreases the odor level. The results of a laboratory odor study are presented in Table 9-3 (7). Concentrations of less than 10 mg/l of powdered activated carbon were successful in providing significant odor reduction.

Table 9-3

Effect of Powdered Activated Carbon on Odor Reduction

Plant	Raw Domestic Wastewater[1] TON	Activated Carbon Treated Effluent TON	Conc. of Activated Carbon mg/l
Charlottesville, Va.	300	100	3.8
Hershey Estates, Pa.	140	100	5.0
Butler, Pa.	280	200	10.0

[1] Sample temperature = 60°C

9.3.3.3 Collection and Treatment of Noxious Gases

The covering of odorous unit process facilities to localize odors is a method which can be used to prevent odors from reaching the atmosphere. The major expenses of this method are the covering of the units and collection and treatment of the evacuated gases. In cold climates, covering units can lead to conditions of high humidity and indoor fog if proper ventilation is not provided. Many municipal plants, e.g., Cedar Rapids, Iowa (8), and Elmira, New York (8), are using low-cost, formed-in-place styrofoam domes on odorous treatment units.

The treatment methods usually considered for evacuated gases include simple or catalytic combustion, ozonation, and chemical oxidation.

Combustion methods require heating the gases to approximately 800°F to 900°F for catalytic combustion and approximately 1,300°F to 1,400°F for simple combustion. Operating costs for these methods are primarily determined by the amount of air to be heated. Ozonation, while somewhat affected by the volume of gas collected, is primarily affected by the quantity of odorous materials to be controlled.

9.3.4 Effects on Subsequent Units

A consideration in using chlorine for odor control is that the chlorine dosage should not produce a high residual chlorine level which may in turn be detrimental to secondary biological units. When using lime for odor control, consideration must also be given to increased sludge production.

9.3.5 Process Designs and Cost Estimates

A cost estimate has been prepared for two odor-control systems (chlorination and powdered activated carbon) for 1, 3, and 5 mgd treatment plants. The capital costs are presented in Table 9-4. The chemicals are added to raw wastewater before the downstream treatment units.

Table 9-4

Capital Costs for Odor Control Systems[1]
ENR Index 1500

Plant Size mgd	Capital Costs for Odor Control Systems	
	Chlorination 10 mg/l	Powdered Activated Carbon 10 mg/l
1	$41,000[2]	$33,000
3	43,000[2]	46,000
5	45,000[2]	64,000

[1] These costs do not include a contingency for engineering design, bonding, and construction supervision.

[2] Smallest size commercially available chlorinator.

The chlorination system included a gas chlorinator capable of delivering 10 mg/l of chlorine during peak flow rates. A building, scale, and other necessary appurtenances were included. The powdered activated carbon system included a 15-day storage hopper and a volumetric feeder capable of delivering 10 mg/l at peak flow rates. In addition, a 1-day capacity slurry tank, pump, building, and associated piping were included.

9.4 Other Uses of Chlorine

In the operation of wastewater treatment plants, chlorine has been found useful as an upgrading technique. Some of the various applications of chlorine are as follows (2) (9) (10):

1. Destruction or control of undesirable growths of algae and slime-forming bacteria in pipelines and conduits.

2. Control of filter flies, clogging, and ponding in trickling filters. Chlorine applied for approximately 8 hours to produce a residual of 1 to 2 mg/l in the distributor arm will generally unclog the filter. Residuals of 20 to 50 mg/l will eliminate ponding by causing the filter to unload all of its biological slime.

3. Improvement in wastewater coagulation.

4. Improvement in the separation of grease from wastewaters.

5. Reduction of the immediate oxygen requirements of return activated sludge and digester supernatant return.

6. Chlorination of return activated sludge may be effective in the control of sludge bulking. A chlorine dose of 1 to 10 mg/l based on the volume of return sludge

has been used, provided the chlorine application point is located to allow for 2 to 3 minutes of mixing before being discharged to the aeration basin.

9.5 References

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2. *Sewage Treatment Plant Design.* Water Pollution Control Federation Manual of Practice No. 8, Washington, D.C., 1959.

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CHAPTER 10

SLUDGE THICKENING

10.1 Air Flotation

The use of air flotation for upgrading is limited primarily to thickening of sludges prior to dewatering. Used in this way, the efficiency and/or capacity of the subsequent dewatering units can be increased and the volume of supernatant from the following digestion units can be decreased. Existing air flotation thickening units can be upgraded by the optimization of process variables, and by the utilization of polyelectrolytes.

Air flotation thickening is best applied to thickening waste activated sludge. With this process, it is possible to thicken the sludge to 6 percent, while the maximum concentration attainable by gravity thickening without chemical addition is 2-3 percent (1). The air flotation process can also be applied to mixtures of primary and waste activated sludge. The greater the ratio of primary sludge to waste activated sludge, the higher the permissible solids loading to the flotation unit. Due to the high operating costs, it is generally recommended that air flotation be considered only for thickening waste activated sludge (2).

10.1.1 Process and Design Considerations

The most commonly used type of air flotation unit is the dissolved air pressure flotation unit. A schematic flow diagram for a typical unit is illustrated in Figure 10-1. In this unit, the recycled flow is pressurized from 40 to 70 psig and then saturated with air in the pressure tank. The pressurized effluent is then mixed with the influent sludge and subsequently released into the flotation tank. The excess dissolved air then separates from solution, which is now under atmospheric pressure, and the minute (average diameter 80 microns) rising gas bubbles attach themselves to particles which form the sludge blanket (3). The thickened blanket is skimmed off and pumped to the downstream sludge handling facilities while the subnatant is returned to the plant.

The following table is a summary of typical parameters used in the design of air flotation thickening units:

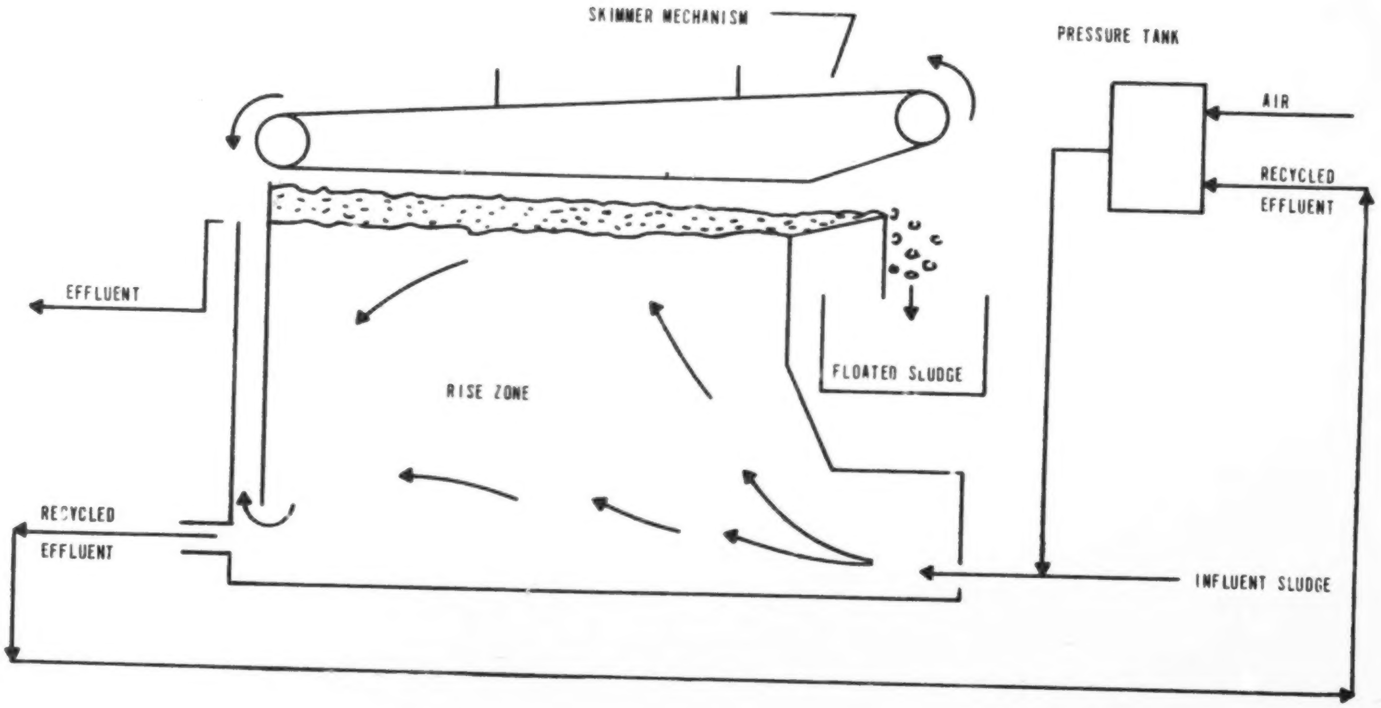
Table 10-1

Air Flotation Parameters

Parameter	Typical Value
Air pressure, psig	40-70
Effluent Recycle Ratio, percent of influent flow	30-150
Detention time, hours	3
Air to solids ratio, lbs air/lb solids	0.02
Solids loading, lbs/sq.ft./day	10-50
Polymer Addition, lbs/ton dry solids	10

FIGURE 10-1

SCHEMATIC OF AN AIR FLOTATION UNIT


SKIMMER MECHANISM
PRESSURE TANK
AIR
RECYCLED EFFLUENT
EFFLUENT
FLOATED SLUDGE
RISE ZONE
RECYCLED EFFLUENT
INFLUENT SLUDGE

In addition to the above parameters, the feed solids concentrations and the type and quality of the sludge affect the performance of the air flotation unit. A detailed discussion of all the previous parameters can be found in the following references (2) (3) (4) (5).

Bench-scale flotation units have been utilized for air flotation designs, but poor correlations have generally been obtained with full-scale performance (3) (5). Therefore, pilot units usually are recommended to determine optimum recycle rates, chemical requirements, and general applicability of air flotation to sludge thickening.

Typical operating data for various air flotation units is presented in Table 10-2. Combined primary and activated sludge produces a more concentrated float sludge than waste activated sludge alone. Polymer and/or chemical addition allows greater solids loading and improves solids recovery without substantially increasing the float solids concentration.

10.1.2 Use of Air Flotation for Upgrading Existing Sludge Handling Facilities

As an upgrading technique, air flotation is best applied to thickening waste activated sludge. The process produces a sludge concentration of 4 to 6 percent, thus decreasing the volume of sludge to be handled in subsequent solids handling units. The capacity and efficiency of an air flotation process can normally be improved with polymer addition of less than 10 pounds/ton of dry solids.

10.1.3 Process Designs and Cost Estimates

The following example illustrates the use of prethickening by air flotation prior to anaerobic digestion.

Existing anaerobic digesters were experiencing unstable operation due to the increased volume of sludge produced by an increase in plant flow from 10 to 16 mgd. Operational data from the overloaded and the upgraded plant are presented in Table 10-3. Prior to upgrading, the waste activated sludge was recycled to the primary clarifier. The volume of the combined sludge was 100,000 gpd at 3 percent solids. The increased sludge volume due to the plant overloading decreased the detention time in the digesters from 17 to 11.25 days. To improve the operation of the existing digesters, it was necessary to reduce the sludge volume to increase the digester detention time. To reduce sludge volume, thickening of the waste activated sludge by air flotation was considered.

Table 10-2

Air Flotation Performance

Description	Solids Loading lbs/sq.ft./day	Influent Suspended Solids percent	Float Solids percent	Solids Recovery percent	Reference
Activated Sludge	12 to 18	0.5 to 1.5	4.0 to 6.0	85 to 95	6
Activated Sludge[1]	24 to 48	0.5 to 1.5	4.0 to 5.0	95 to 99	7
Activated Sludge	13.9	0.81	4.9	85	8
Activated Sludge	7.1	0.77	3.7	99	8
Activated Sludge	19.8	0.45	4.6	83	8
Activated Sludge	26.2	0.80	6.5	93	8
Activated Sludge	28.8	0.46	4.0	88	8
Combined (Primary and Activated) Sludge	24 to 30	1.5 to 3.0	6.0 to 8.0	85 to 95	6
Combined (Primary and Activated) Sludge	21	0.64	8.6	91	8
Combined (Primary and Activated) Sludge	46.6 40.7	2.30 1.77	7.1 5.3	94 88	8 8

[1] 3 to 6 lbs polyelectrolyte/ton.

Table 10-3

Example of
Upgrading Sludge Handling Facilities
Using Air Flotation

Description	Overloaded Plant	Upgraded Plant
Total Primary Solids Produced	15,300 lbs	15,300 lbs
Total Primary Solids	—	36,800 gpd (5%)
Total Waste Activated Sludge Produced	9,800 lbs	9,800
Total Waste Activated Sludge Volume	—	29,400 gpd (4%)
Total Combined Solids Produced	25,100 lbs	25,100 lbs
Total Combined Solids Volume	100,000 gpd (3%)	66,200 gpd (4.5%)
Digester Hydraulic Detention Time	11.25 days	17 days

The flow diagram for the upgraded plant is shown in Figure 10-2 As a result of separate thickening of the waste activated sludge, it is expected that the primary sludge can be concentrated to 5 percent in the primary clarifiers. The total volume of sludge discharged to the anaerobic digestion facility due to the separate thickening of the waste activated sludge and the improved solids concentrations in the primary clarifier is expected to be 66,220 gpd, compared to 100,000 gpd prior to upgrading. The air flotation system was designed using an air pressure of 50 psig, an effluent recycle of 100 percent, and a solids loading of 25 lbs/day/sq.ft. It is anticipated that the polymer dosage requirements will be 5 lbs/ton of dry solids.

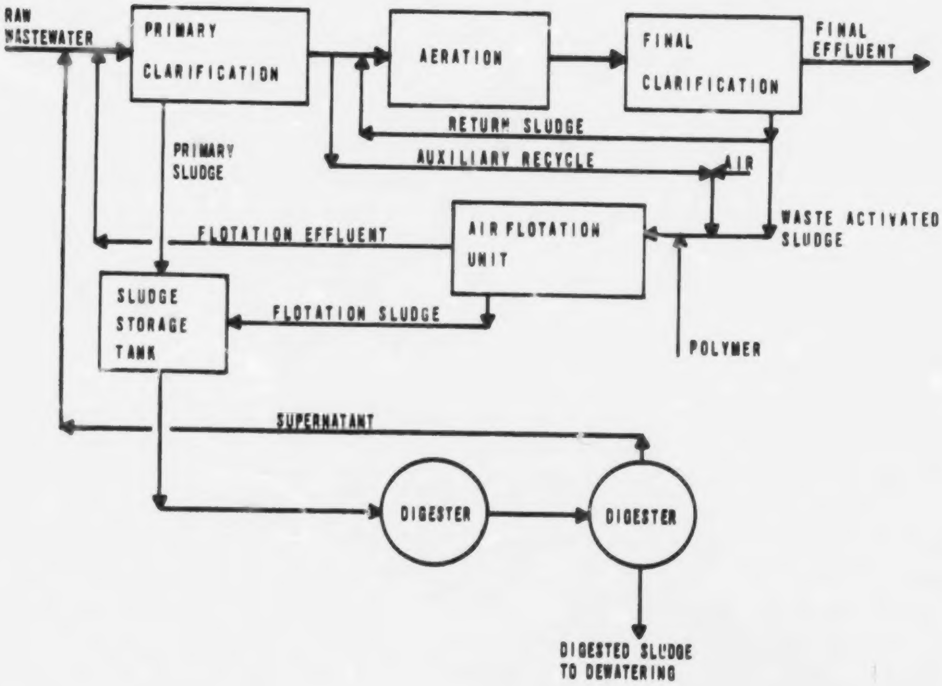
The capital costs for air flotation thickening in this example are estimated at $156,000 (ENR 1500). These costs include two air flotation units, a polymer addition system, and appropriate connecting piping. They do not include an allowance for engineering design, bonding, and construction supervision.

10.2 Gravity Thickening

Gravity thickening is the most common process in use today for the concentration of sludge prior to digestion and/or dewatering. Thickeners can contribute to the upgrading of sludge handling facilities as follows:

1. Increase the capacity of overloaded digesters or subsequent sludge handling units.
2. Reduce the size and increase the efficiency of sludge digestion or dewatering units.
3. Improve primary clarifier performance by providing continuous withdrawal of sludge, thereby insuring maximum removal of solids.

FIGURE 10-2
UPGRADING SLUDGE HANDLING FACILITIES USING AIR FLOTATION


RAW WASTEWATER
PRIMARY CLARIFICATION
AERATION
FINAL CLARIFICATION
FINAL EFFLUENT
RETURN SLUDGE
AUXILIARY RECYCLE
AIR
PRIMARY SLUDGE
FLOTATION EFFLUENT
AIR FLOTATION UNIT
WASTE ACTIVATED SLUDGE
SLUDGE STORAGE TANK
FLOTATION SLUDGE
POLYMER
SUPERNATANT
DIGESTER
DIGESTER
DIGESTED SLUDGE TO DEWATERING

The process is simple and is the least expensive of the available thickening processes. The reduction in size and improvement in efficiency of subsequent sludge handling processes often can offset the cost of gravity thickening. The process also allows equalization and blending of sludges, thereby improving the uniformity of feed solids to the following processes. Existing gravity thickeners can be upgraded by providing continuous feed and drawoff, by diluting the feed solids, and by chemical addition.

10.2.1 Process Considerations

Gravity thickening is characterized by zone settling. The four basic settling zones in a thickener are:

1. The clarification zone at the top containing the relatively clear supernatant.
2. The hindered settling zone where the suspension moves downward at a constant rate and a layer of settled solids begins building from the bottom of the zone.
3. The transition zone characterized by a decreasing solids settling rate.
4. The compression zone where consolidation of sludge results solely from liquid being forced upward around the solids.

To date, many attempts have been made to simulate zone settling in a batch settling test to generate design information which would be applicable to a continuous unit. Various theories have been developed for analyzing batch settling data and they have been reviewed and discussed in the literature (9) (10) (11) (12). Most of the theories assume that the settling velocity of sludge at a given concentration in a small batch cylinder is similar to the velocity in prototype thickening units. However, it has also been recognized that other parameters are involved, such as cylinder depth, cylinder diameter, mixing conditions, and sludge characteristics. All have a definite influence on thickening characteristics. The cumulative effect of these parameters is such that when batch settling test data are used for unit sizing, the result is an oversized unit. For this reason, batch settling test results must be scaled down (10). Edde and Eckenfelder (11) have developed a mathematical model from batch and full-scale thickening data. The model facilitates determination of solids loading at a given sludge blanket depth, initial feed solids, underflow concentration, and hydraulic loading. This technique is particularly useful for determining gravity thickener design parameters when upgrading existing wastewater treatment plants.

The performance of a gravity thickener depends a great deal upon the type of sludge to be thickened. Generally, poor performance is achieved when thickening activated sludge alone. An underflow concentration of 2.0 to 3.0 percent is usually the maximum attainable. Typical performance data for gravity thickeners are presented in Table 10-4.

Table 10-4

Gravity Thickening Process Performance

Type of Sludge	Mass Loading lbs/sq.ft./day	Influent Solids Concentration percent	Effluent Solids Concentration percent	Reference
Primary Sludge	24.2	0.4	4.5	13
	31.2	0.6	4.6	13
	24.9	0.4	4.9	13
	17.2	0.3	4.1	13
	13.2	0.2	3.8	13
	16.3	0.2	4.2	13
Activated Sludge	21.0	1.06	3.0	1
	20.0	0.87	2.8	1
	2.0-3.5[1]	0.20	2.3-2.8	14
	19	0.20	0.74	14
Primary and Secondary Sludge	3.1	0.2	4.5	13
	7.8	0.6	6.3	13
	17.9	1.2	8.1	13
	37.5	0.7	4.0	13
	28.5	0.5	4.5	13
	20.0[2]	1.1	4.4	1
	4.6	1.1	7.8	14

[1] Polyelectrolyte addition.

[2] Ratio 1:1 (primary to secondary sludge), by volume.

10.2.2 Design Considerations

Both solids loading and hydraulic surface loadings must be considered when designing gravity thickeners. Experience indicates that solids or mass loading generally governs the design (15). The following mass loadings (3) (16) have been used for thickener design for different types of sludge:

Table 10-5

Mass Loadings for Designing Thickeners

Type of Sludge	Mass Loading lbs/sq.ft./day
Primary Sludge	22
Primary and Trickling Filter Sludge	15
Primary and Waste Activated Sludge	6-10
Waste Activated Sludge	4-8

The dry solids ratio of waste activated to primary sludge governs the acceptable solids loading to be used in thickener design. As this ratio increases, the acceptable solids loading decreases. In other words, the addition of waste activated sludge reduces the acceptable solids loading to the thickeners.

Most thickeners are operated at a hydraulic loading of 600 to 800 gpd/sq.ft. of surface area (16). Thickeners with hydraulic loadings less than 400 gpd/sq.ft. have been found to produce odors (16). To achieve hydraulic loadings in the acceptable range, secondary effluent is normally blended with the combined waste sludge before feeding the resulting uniform diluted sludge to the thickeners.

As mentioned previously, solids loading is generally the controlling parameter and dictates the required surface area of the thickener. For example, if a solids loading of 10 lbs/sq.ft./day is used for a combined primary and activated sludge and typical performance efficiencies are desired, the calculated hydraulic loading will be in the order of 100 gpd/sq.ft. Effluent dilution will be required to achieve the recommended 600 to 800 gpd/sq.ft.

Most continuous thickeners today are circular and designed with a side water depth of approximately 10 feet. While sludge blanket depth is an important parameter, it has been reported that underflow solids concentrations are independent of sludge blanket depths greater than 3 feet (3).

In the design of gravity thickeners, it is important that operational flexibility be provided. Such flexibility includes the ability to regulate the quantity of dilution water, adequate sludge pumping capacity so that solids concentrations are not limited, continuous feed and underflow pumping, protection devices against torque overload, and a sludge blanket detection device.

10.2.3 Upgrading Existing Facilities with Gravity Thickening

Gravity thickening is generally used prior to digestion processes. It can also be used as a combined thickening and equalization process prior to sludge dewatering. Another application is in areas where sludge hauling is utilized and there is a need to reduce the volume of sludge to be hauled. In all cases, gravity thickening will yield higher underflow solids concentrations than obtainable with primary sedimentation, and the efficiency of subsequent digestion and dewatering facilities will improve. Hence, gravity thickening should always be considered in upgrading existing solids handling and dewatering facilities.

10.2.4 Upgrading Existing Gravity Thickeners

Improved thickening can be obtained by diluting the sludges to be thickened. It has been reported that a feed solids concentration of 0.5 to 1.0 percent is optimum and that dilution reduces the interference between the settling particles (3).

Torpey (17) used dilution for thickening combined primary and secondary sludges in the development of the Densludge System. A feed sludge concentration of less than 1 percent produced underflow concentrations of 11.2 and 6 percent for combined primary and modified waste activated sludge, and combined primary and conventional waste activated sludge, respectively. To obtain these dilute feed sludge concentrations, dilution water was pumped from the primary or secondary clarifier and blended with the combined sludges prior to thickening.

Thickening systems at New York City's Tallmans Island and Bowery Bay pollution control facilities utilize the processes developed by Torpey and presently obtain 4-6 percent underflow solids concentration with a yearly average of 4.5 percent (18). Both plants are operating using a combination of the step and activated aeration processes. A similar design with digested sludge recirculation at Bergen County, New Jersey, produces an underflow concentration of 5.2 to 7.5 percent with a yearly average of 6.3 percent (19). In both cases, the lower underflow concentrations occur during the summer months.

The improved thickening due to dilution can also be attributed to the fines that are washed from the sludge and returned to the plant through the thickener overflow. Experiences at Bergen County, New Jersey, show that, even with digested sludge recirculated to the thickener, the overflow from the thickener does not appreciably affect the over-all BOD removal efficiency of the treatment plant (20).

Similar experiences were reported at the Bowery Bay Pollution Control Plant. However, the air requirements were increased from 0.28 to 0.31 cu.ft./gal. (21) by the return of thickener supernatant to the aeration tanks.

Experiences with the use of polyelectrolytes for upgrading gravity thickening have been reported at Amarillo, Texas (14). The original plant was designed to handle 7.5 mgd, but was receiving a flow of up to 10.5 mgd. A 55-foot diameter gravity thickener was being used to thicken combined primary and waste activated sludge. However, bulking occurred due to the overloaded conditions and the high ratio of primary sludge to waste activated sludge. In-plant recycling of solids resulted.

To minimize the problem, only activated sludge was thickened in the gravity thickener. Polyelectrolyte addition was utilized in the thickener to improve sludge blanket control and to obtain maximum underflow solids concentration. Anionic polyelectrolytes were ineffective, but a cationic polyelectrolyte permitted a solids loading of 4.5 to 7.0 lbs/day/sq.ft. while maintaining an underflow solids concentration of 2.6 percent, at a cost of $1.10 to $3.64/ton of dry solids. Polyelectrolyte was used here to successfully control the sludge blanket height. This practice was continued until the plant was upgraded to 12 mgd (14).

For the expansion to 12 mgd at Amarillo, Texas, an existing 70-foot diameter final clarifier was modified for thickening the waste activated sludge. The overflow from this thickener is mixed with the primary sludge to dilute the feed sludge to the existing primary sludge thickener (14).

Operation of the waste activated sludge thickener showed that, at a solids loading of 2 to 3.5 lbs/sq.ft./day, only a 2.4 percent underflow solids concentration could be obtained. Polymer addition was tried once again to increase the underflow solids concentration during a 142-day program, but proved unsuccessful.

In Chicago, the addition of polymer at dosages of less than 10 lbs/ton dry solids increased the solids loading by 2 to 4 times, but there was no benefit in solids thickening (1). From these two examples, it appears that polymer addition improves solids capture and reduces solids overflow, but has little or no effect on improving solids underflow concentration. The use of pickets has been tried in order to improve thickening and was shown to be successful at Chicago (1), but was unsuccessful at Amarillo, Texas (14).

10.2.5 Process Designs and Cost Estimates

The following example illustrates the use of gravity thickening before anaerobic digestion.

Existing anaerobic digesters were experiencing unstable operation due to the increased amount of sludge generated by an increase in plant flow from 10 to 16 mgd. Operational data from the overloaded plant without gravity thickening and the upgraded plant with thickening are presented in Table 10-6. The volume of the combined primary and secondary sludge was 100,000 gpd at 3 percent solids prior to upgrading. The increased sludge volume due to the plant overloading decreased the detention time in the digesters from 17 to 11.25 days.

Table 10-6

Example of
Upgrading Sludge Handling Facilities
Using Gravity Thickening

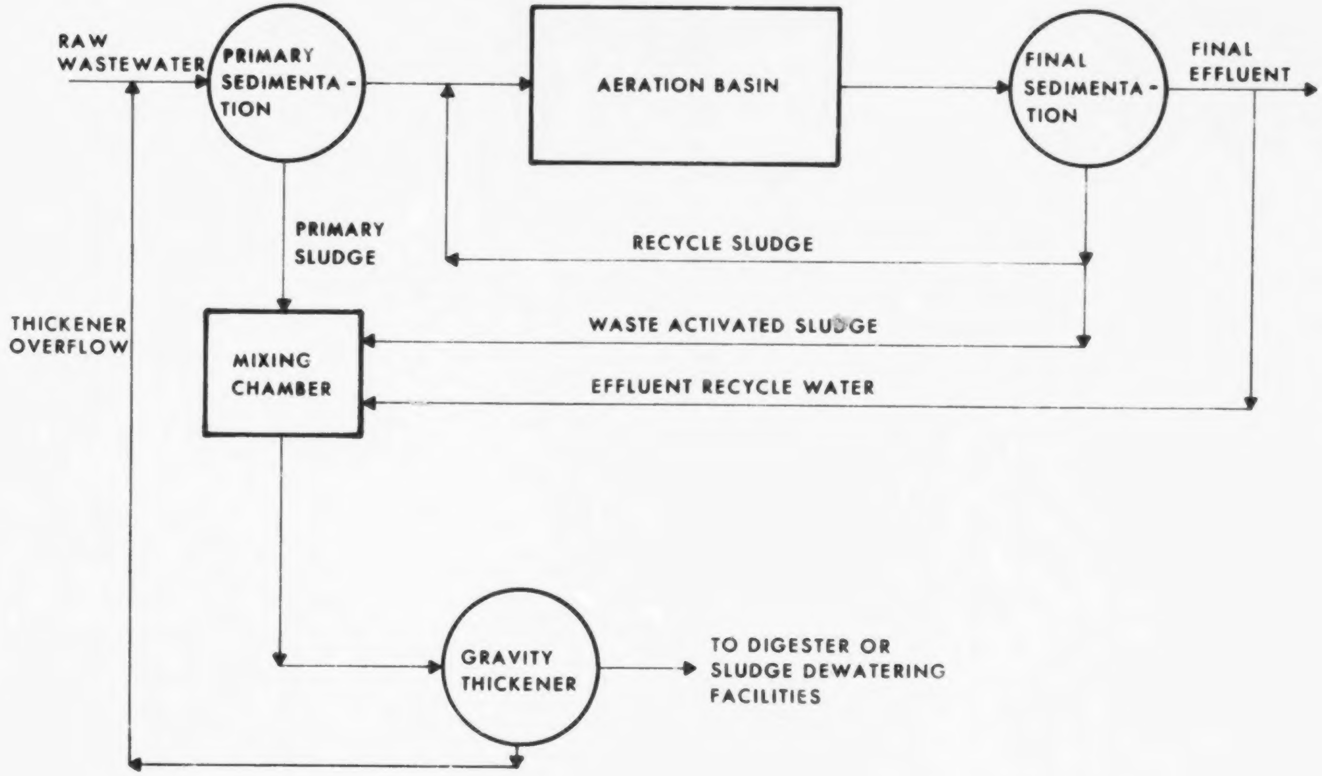
Description	Overloaded Plant	Upgraded Plant
Loadings to the Anaerobic Digester		
Total Primary & Secondary Solids	25,100 lbs	25,100 lbs
Total Volume Primary & Secondary Sludge	100,000 gpd (3%)	66,200 gpd (4.5%)
Digester Hydraulic Detention Time	11.25 days	17 days

To improve the operation of the existing digesters, it was decided to reduce the sludge volume by gravity thickening. The flow diagram of the upgraded plant is shown in Figure 10-3. The total volume of sludge discharged to the anaerobic digestion facility as a result of gravity thickening was reduced from 100,000 gpd at 3 percent solids to 66,200 gpd at 4.5 percent solids. The gravity thickener was designed using a mass loading of 10 lbs/sq.ft./day. A hydraulic overflow rate of 600 gpd/sq.ft. was achieved by recycling final effluent to the mixing chamber ahead of the thickener as shown in Figure 10-3.

The capital costs for gravity thickening in this example are estimated at $258,000 (ENR 1500). These costs include one gravity thickener, a mixing chamber, effluent recycle capacity, and an allowance for appropriate connecting piping. They do not include the allowance for engineering design, bonding, and construction supervision.

FIGURE 10-3

UPGRADING SLUDGE HANDLING FACILITIES USING A GRAVITY THICKENER


RAW WASTEWATER
PRIMARY SEDIMENTATION
AERATION BASIN
FINAL SEDIMENTATION
FINAL EFFLUENT
PRIMARY SLUDGE
RECYCLE SLUDGE
THICKENER OVERFLOW
MIXING CHAMBER
WASTE ACTIVATED SLUDGE
EFFLUENT RECYCLE WATER
GRAVITY THICKENER
TO DIGESTER OR SLUDGE DEWATERING FACILITIES

10.3 References

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CHAPTER 11

SLUDGE DIGESTION

11.1 Anaerobic Digestion

Anaerobic digestion is one of the most frequently employed processes for sludge treatment. The process converts the biodegradable portion of the sludge solids to inoffensive gases. In contrast to raw sludge, which is difficult to dewater, offensive to the senses, and laden with pathogenic organisms, the residue after digestion (digested sludge) is relatively easy to dewater, non-offensive, and contains few pathogens. Thus, anaerobic digestion achieves ultimate disposal by gasifying a portion of the sludge and by preparing the remainder for ultimate disposal by other methods. In addition, the end products have recycle potential. The major gaseous end product is methane, which is often used as a source of fuel in wastewater treatment plants. The digested sludge is an excellent soil conditioner and has found some utility for this purpose.

11.1.1 Biochemical Theory

Operation, control, and design of this process require an understanding of the fundamental biochemistry and bacteriology involved. Consequently, a brief review of these will be given here.

Anaerobic digestion of sludge is a complex biochemical process employing several groups of anaerobic and facultative organisms. In general, the process can be considered to consist of two steps. In the first step, facultative organisms called "acid formers" degrade the complex organics of wastewater sludge to volatile organic acids. Acetic acid is the primary acid formed, with propionic and butyric acids of secondary importance. In the second step, these volatile acids are fermented to methane and carbon dioxide by a group of strict anaerobes called "methane bacteria."

The more important of these two phases is the methane fermentation phase because:

1. The only mechanism of COD or BOD removal is the production of methane. Acid production only solubilizes the complex organics; it does not accomplish stabilization.

2. This step has been found to be the rate-limiting step in the reaction sequence.

The primary reason why the methane fermentation step is rate limiting is that the reproduction rate for these organisms is quite low relative to that of other groups of bacteria. For example, the doubling time of the acid formers is several hours while that of the methane formers is, under ideal conditions, four days. Thus, even if a temporary difficulty in the system arose, it would be much harder for the methane organisms to adjust than for the acid formers. In addition, it has been found that the environmental conditions required to maintain optimum performance by the methane organisms are much more restrictive than for the acid forming organisms. Consequently, most of the effort in design and operation should be expended to make sure that the methane fermentation step is carried out as efficiently as possible.

11.1.2 Environmental Conditions for Optimum Performance

Knowledge of the range of environmental conditions which favor optimum performance of this process is not as extensive as desired. For this reason, the rate of failure with this process is higher than for other waste treatment processes. A summary of the state of knowledge is given below.

11.1.2.1 pH

Tight pH control is required for this process because methane bacteria are extremely sensitive to slight changes in pH. The usual pH range required is 6.6 to 7.4. In general, it is wise to maintain the pH as close to 7.0 as possible.

In an anaerobic digester, pH is maintained by a bicarbonate buffer system due to the great quantity of carbon dioxide produced during methane fermentation. The pH is a function of the bicarbonate alkalinity of the digesting liquor and the fraction of CO_2 in the digester gas. Figure 11-1 prepared by McCarty (1) illustrates this relationship. Because of the significance of pH control in digester operation, it is most important that the dynamic nature of buffer destruction and formation in the digester be understood. This process is reviewed in the following equations for simple carbohydrates such as glucose. The equations mentioned are equally applicable to digestion of sludge.

$$C_6H_{12}O_6 \xrightarrow{\text{acid formers}} 3\ CH_3COOH$$

$$3\ CH_3COOH + 3\ NH_4HCO_3 \longrightarrow 3\ CH_3COONH_4 + 3\ H_2O + 3\ CO_2$$

$$3\ CH_3COONH_4 + 3H_2O \xrightarrow{\text{methane bacteria}} 3\ CH_4 + 3\ NH_4HCO_3$$

The first equation represents the breakdown of glucose to acetic acid by acid forming bacteria. The acid is neutralized, as shown in the second equation, by the biocarbonate buffer. If sufficient buffer is not present, the pH would drop, and the conversion of acetate to methane, as shown in the third equation, would be inhibited. During the third reaction, the buffer consumed in the second reaction is reformed.

In the digestion process, a dynamic equilibrium between buffer formation and destruction is maintained when the process is proceeding satisfactorily. However, when an upset occurs, it is usually the methane bacteria rather than the acid formers which are adversely affected. Therefore, net buffer consumption takes place, and the process is in danger of pH failure. When this occurs, an external source of alkalinity must be added to maintain the pH in the proper range.

FIGURE 11-1

RELATIONSHIP BETWEEN pH AND BICARBONATE CONCENTRATION (1)

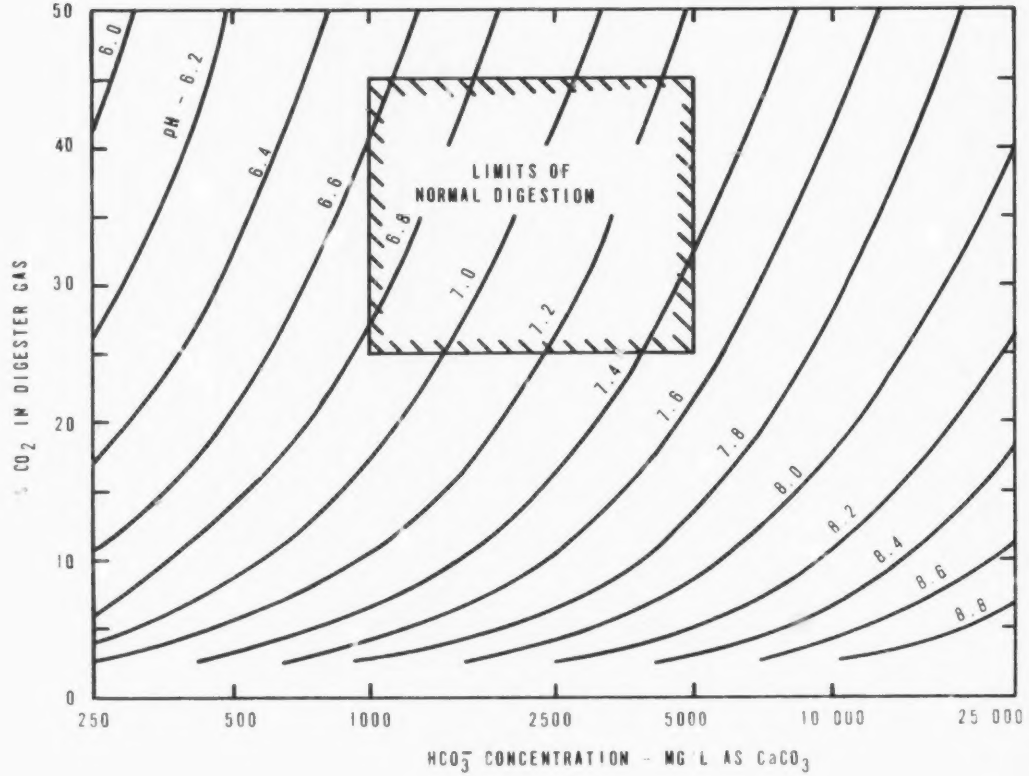


Figure 11-1 indicates that the bicarbonate alkalinity should be maintained at a minimum level of 1,000 mg/l as $CaCO_3$ to ensure adequate pH control. The conventional titration procedure for digester alkalinity determination does not discriminate between the various forms of alkalinity, which is unfortunate because only bicarbonate alkalinity buffers in the correct pH range for good digestion. The other major form of alkalinity measured by this test is that produced by the volatile acids. To determine the bicarbonate alkalinity, both the volatile acid concentration and the total alkalinity must be measured. Then,

$$\text{Bicarbonate Alkalinity} = (\text{Total Alkalinity} - 0.8 \text{ Volatile Acids})$$

The 0.8 factor in the above equation is required to convert the volatile acid units from mg/l as acetic acid to mg/l as $CaCO_3$, the equivalent alkalinity unit.

It should be noted that in the second and third equations, ammonium bicarbonate was used as the form of the alkalinity. This represents the situation in wastewater sludge digestion where large concentrations of ammonium result from the destruction of protein. In fact, the maximum value of the total alkalinity is set by the concentration of the ammonium ion. The carbon dioxide generated in the methane fermentation will not form negatively-charged bicarbonate (the buffer) unless an equivalent quantity of cation is present. This is provided by the destruction of natural protein with the release of positively-charged ammonium. If a cation is not present to force formation of bicarbonate buffer, self-regulation of pH in the digestion process is not possible. In this case, alkaline material must be added continuously to control the pH. For example, the anaerobic degradation of glucose, illustrated in the first three equations, would require the addition of an external source of alkalinity. It is not necessary that ammonium bicarbonate be utilized for this purpose; in fact, in terms of cost and avoidance of potential toxicity, another bicarbonate salt might be favored.

In general, this difficulty will not be experienced in wastewater sludge digestion unless either a high carbohydrate fraction from an industrial waste is present in the sludge or a very thin sludge is being treated.

11.1.2.2 Temperature

The temperature response of methane bacteria is the same as other bacterial groups. Although thermophilic methane bacteria exist, it is generally not economically feasible to heat sludge to this temperature range. Thus, digestion of wastewater sludge is conducted in the mesophilic range. The optimum temperature in this range is 35°C (95°F). More important than maintenance of a particular temperature is maintenance of the chosen temperature at a constant level. A temperature change of 2 or 3 degrees F is sufficient to disturb the dynamic balance between the acid formers and the methane formers. This will lead to an upset because the acid formers will respond much more rapidly to changes in temperature than will the methane bacteria.

11.1.2.3 Nutrients

The major stumbling block in the application of anaerobic treatment to industrial wastewaters is the lack of knowledge of the nutritional requirements of the methane bacteria. Speece and McCarty (2) have done the most definitive work on the macro-nutrient and micro-nutrient requirements of these organisms. As these authors indicate, domestic wastewater appears to contain all of the nutrients required by the methane organisms. Thus, difficulty can be expected in digestion only when a considerable fraction of the sludge is of industrial origin.

11.1.2.4 Toxic Materials

A review of this subject has been provided by Kugelman and Chin (3). These authors indicate that toxicity in general is due to an excess quantity of any material, even for a substance normally considered a nutrient. It was also indicated that a quantitative definition of the concentration at which a substance starts to exert a toxic effect is difficult to define because this could be modified by antagonism, synergism, and acclimation. In addition, the degree of stress on the process as defined by the organic loading and biological solids retention time can significantly affect "toxicity."

Substances which may be present in municipal sludge in concentration ranges which can produce toxicity include heavy metals, sulfides, surface-active agents, light metals, and certain organics. All of these can gain entrance to wastewater sludge from industrial sources. In addition, light-metal cations will enter sludge if an alkaline material is added to control the pH. Several papers (3) (4) (5) review the best engineering data available on toxicity. Reference should be made to these papers for complete information. General information on some substances is given in Table 11-1.

Table 11-1

Concentrations Which Will
Casue a Toxic Situation
in Wastewater Sludge Digestion

Substance	Concentration mg/l
Sulfides	200
Heavy Metals[1]	>1
Sodium	5,000- 8,000
Potassium	4,000-10,000
Calcium	2,000- 6,000
Magnesium	1,200- 3,500
Ammonium	1,700- 4,000
Free Ammonia	150

[1] Soluble

It must be emphasized that the values in this table are only guides. If toxicity is suspected, a thorough analysis of all the chemical constituents of the sludge must be made before definite conclusions can be drawn. Potential solutions to toxicity problems, other than elimination from the wastewater, should be evaluated in small-scale digesters of the type used in laboratory investigations.

11.1.3 Process Kinetics

Lawrence and McCarty (6) have reviewed the kinetics of anaerobic digestion. They indicated that the overall process kinetics are controlled by the methane bacteria. In addition, they found that the removal efficiency could be characterized by the equation:

$$E = \frac{S_o - \dfrac{K_s\,(1 + K_d - SRT)}{SRT \cdot K_m - (1 + K_d \cdot SRT)}}{S_o}$$

where:

K_s, K_m, and K_d = Kinetic constants
S_o = Influent substrate concentration
E = Substrate removal efficiency expressed in decimal form
SRT = Biological Solids Retention Time

The engineer and/or plant operator can control only the SRT. Thus, this is the fundamental design and control parameter which must be used. SRT is the biological solids retention time and is analogous to the sludge age parameter used in activated sludge system design. For a digestion system without sludge recycle, the SRT is numerically equal to the HRT (hydraulic retention time). This analysis indicates a fallacy which is prevalent at present in digestion criteria. Digesters are designed at present on one of three criteria: volume per individual served, weight of volatile solids per unit volume of digester per unit time, and HRT. Of these, the only valid criterion is HRT.

Values for the kinetic constants discussed above were determined experimentally by Lawrence and McCarty (7). These values indicate that at 35°C the absolute minimum SRT for anaerobic digestion is 3 to 4 days. This value agrees with the minimum HRT determined by Torpey (8) in field studies. For design purposes, a longer HRT should be utilized to provide a safety factor against upsets and to allow for fluctuations in sludge volume. In addition, it has been shown that in some situations the rate-limiting step is solubilization of grease and/or protein, which requires HRT values longer than four days. Suggested retention times for high-rate digesters are shown and discussed in the following section.

11.1.4 Present Digestion Systems

Prior to a discussion of procedures for upgrading the performance of digestion systems, a description of existing digestion systems will be presented. Figure 11-2 illustrates the two types of digestion systems in use at present.

In the conventional system, also known as a low-rate system, the tank is not mixed and, in some cases, is not heated. Sludge is added at the top and withdrawn at the bottom. Stratification develops in the system due to a lack of mixing. In general, this can be classed as a plug-flow system. This system is rather inefficient by the normal design criteria utilized, as illustrated in Table 11-2 (9). Because of the lack of mixing and consequent stratification, much of the digester volume is wasted, and many operational problems result. In this type of digester, acidification takes place in the top and middle layers while methane fermentation is confined to the lower layers. This leads to areas of low and high pH in the system, which restrict optimum biological activity. Grease breakdown is poor because the grease tends to float to the top of the digester while the methane bacteria are confined to the lower levels. Methane bacteria are removed with the digested sludge and are not recycled to the top, where they are required. During progression from top to bottom of the digestion tank, the sludge is compressed and gradually dewatered. The water separated from the sludge forms the supernatant layer. The supernatant is high in nitrogen, phosphorus, BOD, COD, and suspended solids. It places an additional organic load on the biological treatment section of the plant and recycles excess nitrogen and phosphorus through the plant. Chemicals added for pH control are not dispersed throughout the tank, and their effectiveness is limited.

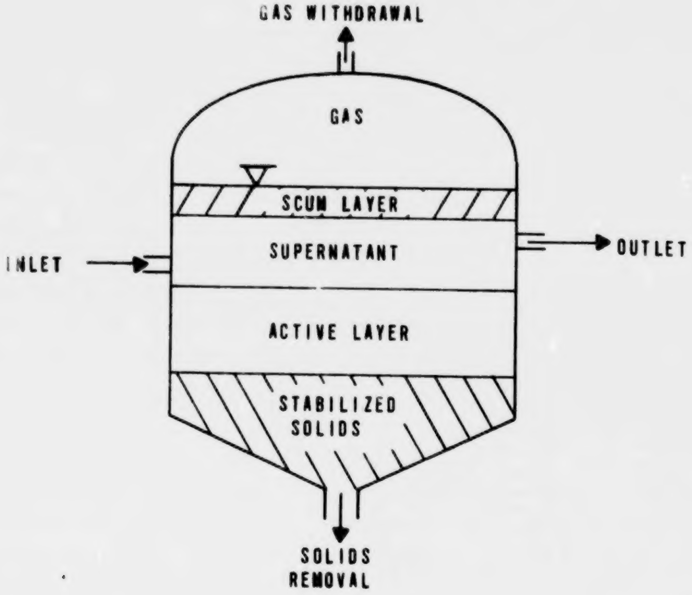
Table 11-2

Typical Design Criteria for Low-Rate and High-Rate Digesters

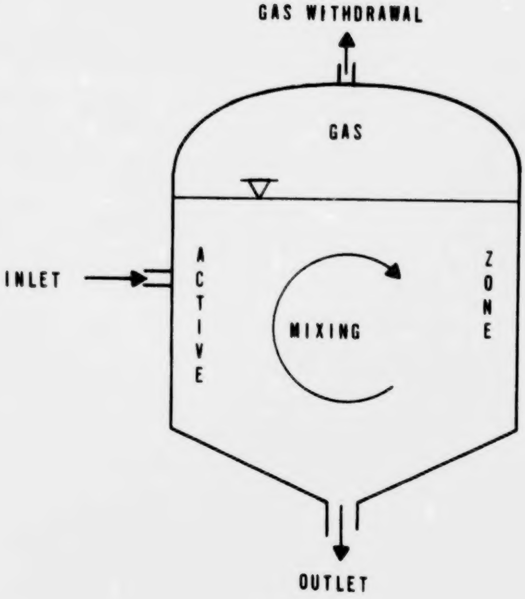
Parameter	Low-Rate	High-Rate
Solids Retention Time (SRT), days	30 to 60	10 to 20
Solids Loading, lb. VSS/cu.ft./day	0.04 to 0.1	0.15 to 0.40
Volume Criteria, cu.ft./capita		
Primary Sludge	2 to 3	1-1/3 to 2
Primary Sludge + Trickling Filter Sludge	4 to 5	2-2/3 to 3-1/3
Primary Sludge + Waste Activated Sludge	4 to 6	2-2/3 to 4
Combined Primary + Waste Biological Sludge Feed Concentration, percent solids (dry basis)	2 to 4	4 to 6
Digester Underflow Concentration, percent solids (dry basis)	4 to 6	4 to 6

Source: Burd (9)

FIGURE 11-2
DIGESTION SYSTEMS


GAS WITHDRAWAL
GAS
SCUM LAYER
SUPERNATANT
INLET
OUTLET
ACTIVE LAYER
STABILIZED SOLIDS
SOLIDS REMOVAL

CONVENTIONAL


GAS WITHDRAWAL
GAS
INLET
ACTIVE
MIXING
ZONE
OUTLET

HIGH RATE

The high-rate system differs from the low-rate system in that the contents are well mixed, either continuously or intermittently, and the digester is heated. This procedure avoids all of the difficulties inherent in low-rate systems. Consequently, this system operates well at lower HRT values and higher organic loading rates. (See Table 11-2).

Various mixing systems have been successfully utilized in digesters. These include:

1. A central draft tube through which sludge is circulated by a turbine mixer set in the table.
2. Propellers (mounted from the roof) which stir the sludge.
3. Gas circulation through diffusers in the base of the digester.
4. Gas injection into the top layer of the sludge.

Specific design of these systems can be obtained from the various manufacturers.

Sludge heating is accomplished either by circulating hot water through coils in the inner wall of the digestion tank or by circulating sludge through an external heat exchanger. The latter method is preferred since it was found the coils inside the digestion tank are easily caked with partially dried sludge. The sludge circulation rate to the external heat exchanger is set to achieve one complete turnover of the tank contents in 24 to 48 hours. Design of a high-rate digestion system must include a heat balance to determine fuel requirements. The WPCF Manual of Practice No. 8 (10) presents in detail the procedure for making such a heat balance. This discussion includes valuable data on the fuel value of sludge gas and the insulation characteristics of typical digester construction material.

One difficulty with high-rate digestion is that the sludge leaving the digester is thinner than the incoming sludge (due to solids destruction). To concentrate the sludge, secondary digesters have been added to many high-rate digestion systems. In effect, these are settling tanks since they are neither heated nor mixed. They also serve as a source of seed sludge in case of digester upset. The secondary digester capacity is usually 2 to 4 times that of the primary digester.

11.1.5 Upgrading Existing Anaerobic Digestion Facilities

Improperly functioning digestion systems can be upgraded by applying procedures which will make the systems more closely approach the theoretical optimum performance. The conditions which will produce optimum performance of this process have been given in the preceding sections of this chapter. Specific upgrading techniques will now be discussed.

11.1.5.1 Process Monitoring and Biochemical Control

The first step in any upgrading technique is constant monitoring of the process for biochemical upset. This can be accomplished with the aid of the volatile acid and alkalinity tests and by a digester gas analysis. Any sudden rise of volatile acids indicates that the system is out of biochemical balance. A rise in the CO_2 fraction in the gas or a decrease in methane production per pound of volatile solids added will also indicate upset. However, the volatile acid test is more sensitive. When an upset occurs, an alkaline material must be added to maintain the bicarbonate alkalinity above 1,000 mg/l as $CaCO_3$. An easily soluble bicarbonate salt, such as $NaHCO_3$, is best for this purpose. Care must be exercised not to exceed the level at which the cation of the alkaline material will cause toxicity. If this is a potential problem, a mixture of alkaline salts should be used. Kugelman and McCarty (4) have described methods of preventing cation toxicity by adding appropriate quantities of cation antagonists.

Control of pH during an upset is only a stop-gap measure. The cause of the upset must be located and eliminated. Sometimes this is easy. For example, heavy-metal toxicity can be completely eliminated by precipitation of the metal in the digester as the sulfide (11). In other cases, only exclusion of the toxin from the system will suffice.

11.1.5.2 Upgrading Techniques

The major upgrading technique for low-rate digesters is conversion to high-rate digestion. To maintain high-rate digestion, the following conditions are necessary:

1. Solids thickening to maintain volatile solids loading in the range of 0.15 to 0.4 lb. VSS/cu.ft./day.
2. Complete mixing of digester contents.
3. Solids feed and withdrawal at a uniform rate.
4. Temperature control system capable of maintaining a uniform temperature range of 30 to 35°C.
5. A solids retention time of 10 to 20 days.

The principal techniques used for upgrading high-rate digesters are to increase feed solids concentration, provide a secondary digester for liquid-solids separation, and increase SRT by recycling digested sludge to the primary digester.

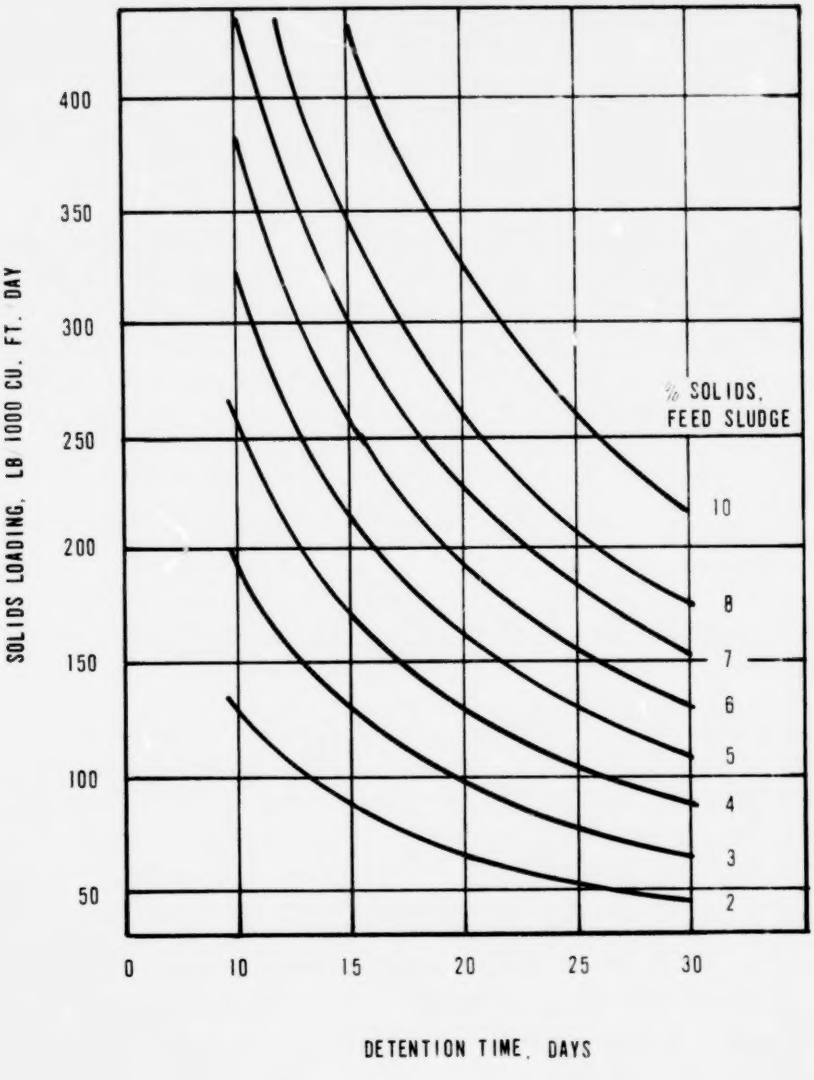
The relationship between thickening of solids, detention time, and solids loading has been illustrated by Sawyer (12) and is shown in Figure 11-3. This relationship points out the importance of thickening the solids prior to digestion.

In most cases, thickening is best accomplished in a separate thickening unit. (Thickening techniques are discussed in Chapter 10 of this manual.) If air flotation is chosen, consideration should be given to minimizing the bound air in the float sludge before pumping it to the digester.

Thickening can also be accomplished in a clarifier and controlled by a sludge density meter as reported by Garrison, (13) and Sironen and Lee (14), or by using automatic

FIGURE 11-3

RELATIONSHIPS BETWEEN SLUDGE SOLIDS, DIGESTER LOADINGS, AND DETENTION TIME (12)*



*OPTIMUM TEMPERATURE RANGE 85-95°F

sludge blanket control devices. Both of these techniques require some visual operator control, as described in the WPCF Manual of Practice No. 16 (15), and will most likely reduce the efficiency of the clarification system.

Pre-thickening is generally required when solids loadings greater than 0.2 lb.VSS/cu.ft./day are used. At these levels, liquid-solids separation in the digestion system becomes more difficult (as previously described), especially if secondary digester capacity is limiting. Hence, if supernatant is drawn from this process, it could be detrimental to the efficiency of the secondary biological system unless properly distributed.

There are restrictions on the degree to which raw sludge can be thickened. These include the difficulty of pumping thick sludge and the maintenance of adequate mixing in the digester. Generally, sludge can be thickened to about 8 percent solids without the need to install special pumping or additional mixing equipment. If it is desired to thicken beyond this point, adequate studies of the sludge flow characteristics must be made to evaluate pumping and mixing requirements.

Potential toxic effects may also limit the degree of sludge thickening. Thickening will result in high ammonium and bicarbonate concentrations in the digester. The higher alkalinity will tend to raise the pH and convert the ammonium ion to free ammonia according to the following equation, raising the possibility of free ammonia toxicity:

$$NH_4^+ \rightleftharpoons H^+ + NH_3$$

If the pH is maintained below 7.2, free ammonia toxicity can generally be avoided. Thus, when digesting thickened sludge, process control is extremely important. If the ammonium ion concentration and pH are high, it may be necessary to add acid to keep the pH below 7.2. The only acid which should be used for this purpose is hydrochloric acid. Sulfuric acid addition would yield sulfate, which would eventually be reduced to corrosive H_2S under anaerobic conditions. Nitric acid addition would release nitrate, which under certain conditions is toxic to methane bacteria.

The addition of mixing by itself can have a significant beneficial effect on digester performance. At the City of Pontiac, Michigan (16), modifications were made to an existing digester by adding a gas recirculation unit to improve mixing. Mixing inhibited scum formation, improved heat transfer, and provided a more stable digestion process. Many other communities have had similar success (17).

Experiences at Chicago (18) have shown that digesters can be upgraded to operate at volatile solids loadings of 0.2 lb./cu.ft./day and at solids retention times of 10 days. Complete mixing is necessary to achieve these operating results and enables a wide variation in loadings.

Bergen County, New Jersey, is the best reported example of upgrading low-rate digesters to high-rate digesters (19). In 1951, a 20-mgd activated sludge plant was constructed with four 1.3-million gallon digesters. In 1961, the capacity of the plant was increased to 50 mgd, and the existing digesters were modified to high-rate. In 1969, additional studies were conducted on a full-scale basis, and it was found that two of the original digesters could handle the entire loading from the existing 50-mgd activated sludge plant. On this basis, it has been projected that the original four digesters will be able to handle the increased solids loading when the plant is expanded to 75 mgd. It is significant to note that even at these higher loadings, the anaerobic process was very stable and the efficiency of the process remained the same.

Successful upgrading of existing digestion facilities at Bergen County was accomplished by the following methods:

1. Completely mixing the digester contents.
2. Pre-thickening the primary and secondary sludge to an average concentration of 6.3 percent (range 5.2 to 7.5 percent).
3. Increasing the applied solids loading from 0.22 lb./cu.ft./day to 0.5 lb./cu.ft./day as a result of thickening.
4. Decreasing the liquid detention time from 22 to 10 days.

The addition of a second stage to a high-rate digester enables digestion in the primary stage and solids-liquid phase separation to occur in the secondary stage. In some areas, this phase separation may be important. It appears that, as the VSS loading in the primary digester increases, the detention time in the secondary sludge digester should be increased.

The loading to two-stage high-rate digestion systems may be increased by recirculating the digested sludge from the secondary stage back to the primary stage, since this effectively increases the SRT at the same digester HRT. As with activated sludge systems, the limiting factor is the solids-liquid phase separation. Perhaps a degasification process can be included between the primary and secondary tanks to aid in the separation, as is done with the anaerobic contact process (20). When satisfactory phase separation is obtained in the secondary digester, adequate SRT's can be maintained while decreasing hydraulic retention times to less than 10 days (21). A more stable digestion process then results, due to higher populations of methane bacteria in the primary stage and to the lessening of toxicity effects at longer SRT's.

Digesters can also be upgraded by recirculating a portion of the digested sludge back to the thickening units and mixing it with the incoming combined sludge and effluent recycle water. This procedure has been reported by Torpey and Melbinger (22) with modified aeration plants in New York City, and was originally adopted to improve the pumping characteristics of highly concentrated digester feed sludge of 10 to 14 percent solids. It appears that the thickening process was improved because the digested sludge was incorporated into the pore spaces of the mixed primary and waste activated sludge, thereby eliminating the typical gel structure produced by grease in the raw sludge. It was also found that recycling digested sludge improved digester performance due to the seeding of the combined sludge prior to digestion and because of the greater SRT thereby afforded. Volatile solids reduction was also increased. A digested sludge recycle of 50 percent appeared to be optimum for the New York City plants, and a net volume reduction of digested sludge from 197 cu.ft./million gallons to 112 cu.ft./million gallons was achieved.

Fines are washed out in the thickener by the high volume of effluent recycle water used and are returned to the aeration basins. This thickener supernatant return increased the aeration requirements of the New York City plants slightly, but had little apparent effect on BOD removal (average of 68 percent removal with recirculation and an average 63 percent without recirculation). However, when efficiencies greater than 90 percent BOD removal are required, thickener supernatant return will definitely be a major concern.

It must be pointed out that conversion to high-rate digestion is not a cure-all, especially if digested sludge is to be dewatered prior to final disposal. Methane production and volatile solids reductions are approximately the same at high-rate as at standard-rate, but indications are that dewatering of high-rate sludge is difficult (9). To obviate this difficulty, Sawyer (12) has suggested that secondary digesters in high-rate digestion systems be two to four times the capacity of high-rate primary digesters, to provide adequate time for solids conditioning. This technique has been reported at Grand Rapids, Michigan (23), where the volatile solids loading to the primary digester is in excess of 0.25 lb./cu.ft./day. The ratio between secondary and primary digester capacity is 3.5:1. Secondary digester underflow solids exceed 10 percent and supernatant solids average less than 2 percent of the raw solids load. This indicates that the economics of decreasing the detention time in the primary digesters should be weighed against providing the additional capacity in the secondary digesters when solids dewatering is required. Such an evaluation would not be required if the digested sludge is to be disposed of on land, because the degree of sludge conditioning is not the same as that required for sludge dewatering.

11.1.6 Anaerobic Supernatant Treatment

The return of supernatant liquor from digesters or thickeners to the treatment facilities is an important consideration. Such supernatants contain a significant quantity of volatile solids, organic matter, and high concentrations of nutrients, particularly nitrogen and phosphorus, as indicated in Table 11-3. The supernatant return problem can be reduced significantly by treating the supernatant with the lime precipitation process followed by ammonia stripping, as shown schematically in Figure 11-4. Operational data for this process are shown in Table 11-3 for a lime concentration of 6,000 mg/l (24). The data indicate substantial reductions in nitrogen, phosphorus, organics, and solids.

FIGURE 11-4
LIME PRECIPITATION PROCESS FOR ANAEROBIC
DIGESTER SUPERNATANT (24)

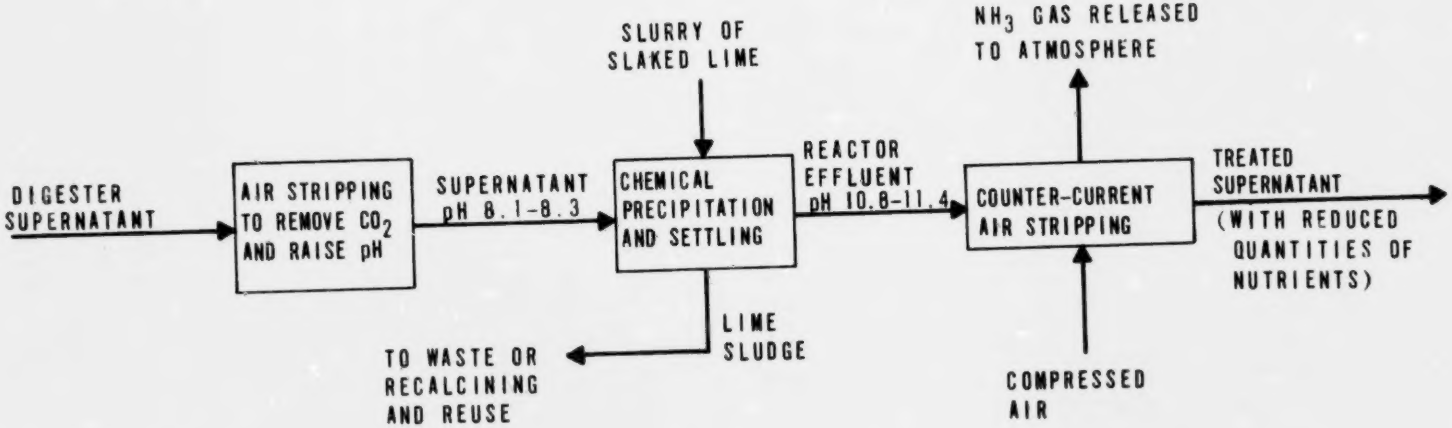

DIGESTER SUPERNATANT
AIR STRIPPING TO REMOVE CO_2 AND RAISE pH
SUPERNATANT pH 8.1-8.3
SLURRY OF SLAKED LIME
CHEMICAL PRECIPITATION AND SETTLING
LIME SLUDGE
TO WASTE OR RECALCINING AND REUSE
REACTOR EFFLUENT pH 10.8-11.4
NH_3 GAS RELEASED TO ATMOSPHERE
COUNTER-CURRENT AIR STRIPPING
COMPRESSED AIR
TREATED SUPERNATANT (WITH REDUCED QUANTITIES OF NUTRIENTS)

Table 11-3

Operational Data for the Lime Precipitation Process for Anaerobic Digester Supernatant

Parameter	Concentration, mg/l	
	Influent	Effluent
pH	7.1	10.7
Total Solids	4,985	2,753
Total Volatile Solids	3,330	1,821
Suspended Solids	2,905	1,190
Volatile Suspended Solids	2,530	930
COD	5,407	2,919
Total Carbon	3,075	1,214
Total Organic Carbon	1,624	914
Ortho - PO_4 (as P)	91	5.9
Total Phosphate (as P)	141	37
NH_3-Nitrogen (as N)	818	726[1]
	--	157[2]
Organic Nitrogen (as N)	282	176

[1] Effluent not air stripped after the lime treatment.
[2] Effluent air stripped after the lime treatment.

Source: Bennett (24)

11.1.7 Process Designs and Cost Estimates

Two examples of upgrading existing anaerobic digesters are presented in this section.

11.1.7.1 Example A

In this example, upgrading of two-stage low-rate digestion facilities was required due to the increase in plant flow from 1 mgd to 3 mgd. A gravity thickener was added prior to digestion. Primary digester performance was improved by adding gas mixing and installing external heat exchangers to control the temperature more accurately. The comparison between existing and upgraded design conditions is presented in Table 11-4. The flow diagram of the upgraded plant is shown in Figure 11-5.

A mixing chamber ensures proper blending of sludges and effluent water prior to thickening. With thickening, the solids concentration and the volatile solids loading to the primary digester were increased from 2 to 5 percent and from 0.036 to 0.108 lb./cu.ft./day, respectively.

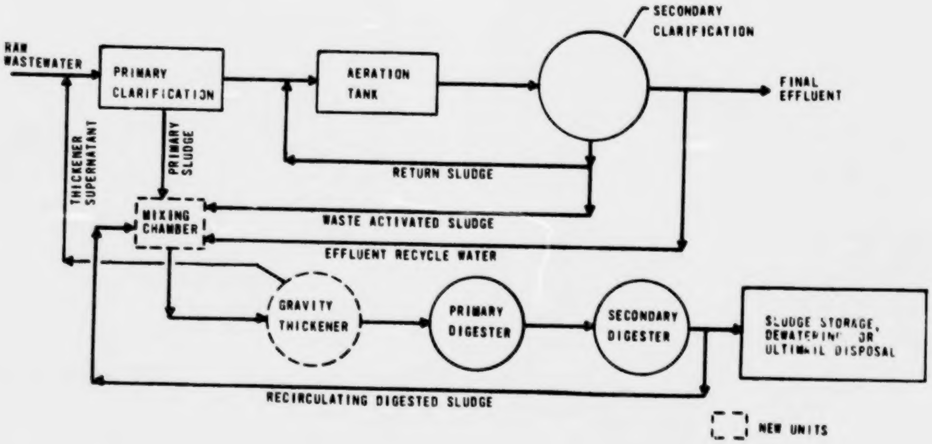
Table 11-4

Upgrading an Existing Low-Rate Digestion System Using Pre-thickening of the Combined Sludge and Improvements to the Primary Digester - Example A

Parameter	Original Design	Upgraded Design
Plant Flow, mgd	1	3
Combined Sludge Characteristics		
Volume, gpd	9,100 (2%)	27,300 (2%)
Solids Contribution, lbs/day	1,530	4,590
% VSS	70	70
Gravity Thickener		
Number	–	1
Solids Loading, lbs./sq.ft./day	–	10
Effluent Water Required to Dilute Sludge, gpd	–	248,100
Hydraulic Loading, gpd/sq.ft.	–	600
Thickened Sludge Volume, gpd	–	10,920 (5%)
Digester		
Number - Primary Digesters	1	1
Number - Secondary Digesters	1	1
Primary Digester Characteristics	Limited heating and mixing	New gas mixing and improved heating
Secondary Digester Characteristics	No heating or mixing	No heating or mixing
Digester Volume Allocation, cu.ft./capita/day	6 (Low-rate)	2 (High-rate)
Digester Volume (Total), cu.ft.	60,000	60,000
Hydraulic Retention Time, days (Total)	49.4	41.0
days (Primary)	24.7	20.5
VSS Loading, lb./cu.ft./day (Total)	0.018	0.054
lb./cu.ft./day (Primary)	0.036	0.108

FIGURE 11-5, EXAMPLE A

UPGRADING AN EXISTING LOW RATE DIGESTION SYSTEM USING PRE-THICKENING OF THE COMBINED SLUDGE AND IMPROVEMENTS TO THE PRIMARY DIGESTER


RAW WASTEWATER
PRIMARY CLARIFICATION
AERATION TANK
SECONDARY CLARIFICATION
FINAL EFFLUENT
THICKENER SUPERNATANT
PRIMARY SLUDGE
RETURN SLUDGE
MIXING CHAMBER
WASTE ACTIVATED SLUDGE
EFFLUENT RECYCLE WATER
GRAVITY THICKENER
PRIMARY DIGESTER
SECONDARY DIGESTER
SLUDGE STORAGE, DEWATERING OR ULTIMATE DISPOSAL
RECIRCULATING DIGESTED SLUDGE
NEW UNITS

Digested sludge recirculation was utilized, with provisions for recirculating 50 percent of the volume of sludge to be digested back to the thickener. Torpey found that this technique improved VSS reduction (22). The capital costs of this upgrading were estimated at \$118,000[1] (ENR 1500) and were allocated as follows:

Thickener	\$ 64,000
Digester Renovation	54,000
TOTAL	\$118,000

The cost of this upgrading is estimated at \$39,500/1,000 lbs./day of increased solids loading.

11.1.7.2 Example B

This example illustrates upgrading of existing digesters to increase capacity by converting both low-rate primary and secondary digesters to high-rate digesters. This is illustrated in Figure 11-6. It is noted that the existing four digesters have all been upgraded, but that the fourth digester is normally used as a storage tank and a backup digester only during periods of operational problems. All upgraded digesters are provided with complete mixing, uniform solids feeding and withdrawal, and uniform temperature control. The design parameters for the existing and upgraded digesters are shown in Table 11-5.

Generally, some form of pre-thickening is required to obtain 4 percent solids in the feed sludge. However, experience at Grand Rapids, Michigan (25) has shown that concentrations as high as 7 percent can be obtained with very close operational control of the primary clarifiers. Therefore, pre-thickening of sludge is assumed not to be necessary in this example.

With this conversion, it must be recognized that there may be problems with the dewatering characteristics of the sludge. This modification will be useful where ultimate disposal land sites can accept 5 to 7 percent digested sludge. The cost associated with this modification is \$92,000, or \$3,000/1,000 lbs./day of increased solids loading. The greatly reduced cost of upgrading in Example B compared to Example A is due to availability of mixing and heating equipment in the existing digesters.

11.2 Aerobic Digestion

Aerobic digestion is an effective means for upgrading existing overloaded sludge digestion facilities, particularly as applied to the digestion of waste activated sludges. It offers a low capital cost means of achieving stabilization and volume reduction of wastewater sludges. Most package type activated sludge plants include aerobic digestion. The process

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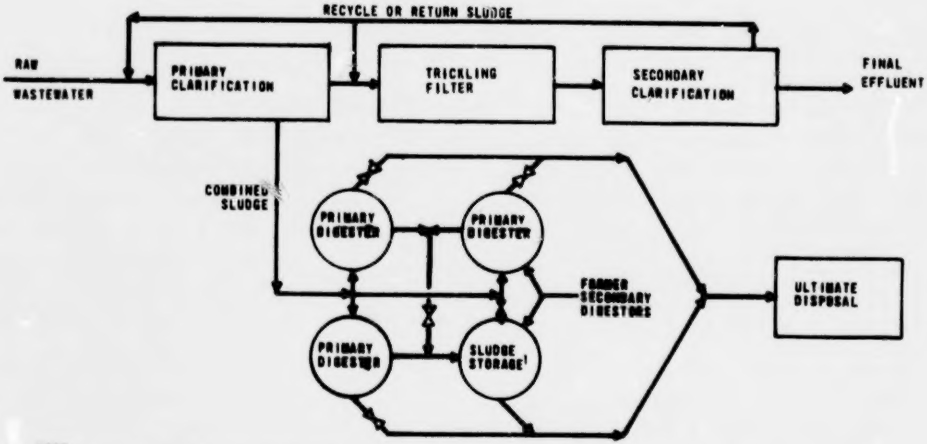
[1] These costs do not include a contingency for engineering design, bonding, and construction supervision.

FIGURE 11-6, EXAMPLE B
UPGRADING EXISTING TWO-STAGE DIGESTERS
TO PRIMARY HIGH-RATE DIGESTERS


RECYCLE OR RETURN SLUDGE
RAW
WASTEWATER
PRIMARY
CLARIFICATION
TRICKLING
FILTER
SECONDARY
CLARIFICATION
FINAL
EFFLUENT
COMBINED
SLUDGE
PRIMARY
DIGESTER
PRIMARY
DIGESTER
PRIMARY
DIGESTER
SLUDGE
STORAGE¹
FORMER
SECONDARY
DIGESTORS
ULTIMATE
DISPOSAL

NOTE:
1. SLUDGE STORAGE UNIT TO BE USED AS BACKUP PRIMARY DIGESTER

Table 11-5

Upgrading Existing Two-Stage Digesters to High-Rate Digesters - Example B

Parameter	Original Design	Upgraded Design
Plant Flow, mgd	12.5	37.5
Digester		
Number - Primary Digesters	2	3
Number - Secondary Digesters	2	1 (Storage Unit)
Primary Digester Characteristics	Mixing and heating provided (2 units)	Mixing and heating provided (3 units)
Secondary Digester Characteristics	No mixing or heating	Storage unit and standby primary digester
Digester Volume Allocation, cu.ft./capita/day	6	2
Digester Volume (Total), cu.ft.	300,000	225,000 (Active) 75,000 (Storage)
Sludge Volume to Digesters, gpd	61,000 (3%)	137,000 (4%)
Solids Loading to Digesters, lbs./day	15,200	45,600
% VSS	70	70
Hydraulic Retention Time, days (Total)	36.8	–
days (Primary)	18.4	12.3 (3 units)
VSS Loading, lb./cu.ft./day (Total)	0.035	–
lb./cu.ft./day (Primary)	0.070	0.142 (3 units)

can also be applied to the digestion of primary sludges or to combinations of primary and secondary sludges. It has been indicated that aerobic digestion is competitive with anaerobic digestion for activated sludge plants up to a size of at least 8 mgd (26).

The typical concentrations of various constituents present in aerobic and anaerobic supernatant liquors, shown in Table 11-6 (24) (27), indicate that the effect of supernatant return on biological units would definitely be less pronounced when aerobic digestion is used.

Table 11-6

Comparison of Aerobic and Anaerobic Supernatant Liquors

Parameters	Aerobic Supernatant	Anaerobic Supernatant
pH	5.6	7.1
BOD, mg/l	16	–
COD, mg/l	–	5,407
Volatile Solids, mg/l	39.5	3,330
Ammonia Nitrogen, mg/l	1.75	818

Fewer operational problems are associated with aerobic digestion than with anaerobic digestion. Hence, less laboratory control and daily maintenance are required. Also, the dangers of gas explosions are eliminated because the only gaseous by-products of aerobic stabilization are carbon dioxide and water vapor.

11.2.1 Process Considerations

Aerobic digestion is accomplished by aerating waste sludge until it is stable and relatively nuisance free. In the aerobic digestion of waste sludges, two different forms of oxidation take place, as shown in the following two reactions. First, a portion of the organic substrate in the wastewater sludge is oxidized and the remainder is converted to cell mass. Second, the cell mass produced or present is oxidized until only a relatively inert fraction remains.

$$\text{organic matter} + O_2 \xrightarrow{\text{bacteria}} \text{cellular matter} + CO_2 + H_2O$$

$$\text{cellular matter} + O_2 \xrightarrow[\text{destruction}]{\text{volatile solid}} \text{digested sludge} + CO_2 + H_2O$$

Once the organic substrate is removed from the system, the biological cells must begin to use their own stored cellular material and dead cells as food. This self-oxidation of cell material (endogenous respiration) reduces the amount of cellular material remaining. Approximately 15 days of detention time are required to stabilize waste biological sludges and to reduce the volatile suspended solids (VSS) by 40 to 60 percent (28). The oxygen requirements for the aerobic digestion process, exclusive of nitrification requirements, are in the range of 3 to 4 mg/l/hr./1,000 mg/l MLSS under aeration for the endogenous respiration phase. This is less than the requirements for the oxidation of raw organic matter.

Both pH and alkalinity are reduced in a properly functioning aerobic digester when nitrification takes place. Nitrification lowers pH according to the following reaction:

$$NH_4^+ + 1.5\,O_2 \xrightarrow{\text{bacteria}} NO_2^- + 2H^+ + H_2O$$

The second step of nitrification is as follows:

$$NO_2^- + 0.5\,O_2 \xrightarrow{\text{bacteria}} NO_3^-$$

Theoretically, 7.1 lbs. of $CaCO_3$ alkalinity are destroyed per lb. of ammonia nitrogen oxidized, since the two protons released neutralize one mole of $CaCO_3$ according to the following reaction:

$$2H^+ + CaCO_3 \longrightarrow H_2CO_3 + Ca^{++}$$

The aerobic digester operational data shown in Table 11-7 indicate the relation between ammonia, nitrite, and nitrate nitrogen as a function of detention time (29). Table 11-7 also indicates that, in the normal temperature ranges of 15° to 35°C, an increase in temperature increases the rate of VSS reduction.

Table 11-7

Aerobic Digestion Operational Data

Detention Time	Temperature	VSS Reduction	pH	Alkalinity	NH_3-N	NO_2-N	NO_3-N
days	°C	percent		mg/l	mg/l	mg/l	mg/l
5	15	21	7.6	510	54	Trace	None
10	15	32	7.6	380	3.2	1.28	64
30	15	40.5	6.6	81	4.0	0.36	170
60	15	46	4.6	23	38	0.23	835
5	20	24	7.6	590	54	Trace	None
10	20	41	7.6	390	4.9	0.59	60
15	20	43	7.8	560	7.0	2.27	29
30	20	44	5.4	31	28	0.19	275
60	20	46	5.1	35	7.0	0.51	700
5	35	26	7.9	630	14	0.18	None
10	35	45	8.0	540	10.0	0.08	None

Source: Jaworski (29)

There are two methods of operating aerobic digesters. One is on a continuous basis, and the other is on a modified or intermittent batch basis. Ritter (30) has reported that it is good practice to stop wasting sludge into the digester for a period of up to 5 days prior to making a sludge withdrawal when using an intermittent batch operation. Continuous digester operation requires provision for continuously decanting the supernatant, but offers the advantages of equalizing air requirements and providing a slow continuous supernatant return.

Published data on phosphate removal in aerobic digesters is extremely limited. Recent testing in Pennsylvania indicated that orthophosphate removal was generally negligible (30). The aerobic digester was operated on an intermittent basis with the supernatant periodically decanted. The operating data revealed that, on days when the digester supernatant was decanted, the effluent concentrations of orthophosphate exceeded the influent concentration by as much as 200 percent (43 mg/l compared to 14 mg/l). This was explained by the fact that when aeration is in progress, phosphate is adsorbed by sludge cells. Conversely, when aeration is terminated, the phosphate is released by sludge cells. If low effluent phosphate concentrations are required from aerobic digesters, then automatic supernatant decanting should be used without stopping the aeration, as in the continuous operation described above.

It has been shown that the detention time in an aerobic digester treating waste sludge from a contact stabilization operation should be 15 days, to obtain satisfactory thickening and dewatering (26). However, it has also been reported that satisfactory digestion of waste trickling filter sludge does not occur until after 30 days of digestion (31). These reports illustrate the need for running a small (55-gallon) aerobic digestion pilot plant on an existing wastewater sludge to help evaluate aerobic digestion characteristics prior to installing a plant-scale unit.

11.2.2 Design Basis

In designing an aerobic digestion system, care must be taken to see that the characteristics of the sludge to be digested are fully identified. As previously mentioned, this can best be done with pilot plant studies. Through such studies, the stabilization time, oxygen requirements, and volatile suspended solids reduction can be determined. A procedure for analyzing aerobic digestion kinetics has been reported by Reynolds (32).

Table 11-8 contains a summary of parameters used in the design of aerobic digestion units for municipal wastewater sludges. As previously discussed, when phosphorus removal is a consideration, continuous operation may be necessary. If phosphorus removal is not a design criterion, then the digester may be operated intermittently. When considering aeration requirements, recognition must be given to that needed for oxidation of the ammonia present.

Table 11-8

Aerobic Digestion Design Parameters

Parameter	Value	Remarks	Reference
Detention Time, days	15-20	Waste Activated Sludge Alone	–
	20-25	Primary + Waste Activated Sludge	–
Air Requirements			
diffuser system, cfm/1,000 cu.ft.	20-35[1]	Enough to keep the solids in suspension and maintain a D.O. between 1-2 mg/l.	9,31
cfm/1,000 cu.ft.	>90[2]		9
mechanical system, gp/1,000 cu.ft.	1.0-1.25	This level is governed by mixing requirements. Most mechanical aerators in aerobic digesters require bottom mixers for solids concentration greater than 8,000 mg/l, especially if deep tanks (>12 feet) are used.	
mg O_2/l/hr./1,000 mg/l MLSS	3.0		26
Minimum Dissolved Oxygen, mg/l	1.0		30
Temperature, °C	>15	If sludge temperatures are lower than 15°C, additional detention time should be provided so that stabilization will occur at the lower biological reaction rates.	
Volatile Solids Reduction, percent	40-50		
Tank Design		Aerobic digestion tanks are open and generally require no special heat transfer equipment or insulation. For small treatment systems (0.1 mgd), the tank design should be flexible enough so that the digester tank can also act as a sludge thickening unit. If thickening is to be utilized in the aeration tank, sock-type diffusers should be used to minimize clogging.	9
Power Costs			
$/yr./lb. BOD removed	2.18	These cost data are based upon three operational plants in Pennsylvania.	30
$/yr./capita	0.37		

[1] Waste activated sludge alone.
[2] Primary and waste activated sludge.

11.2.3 Use of Aerobic Digestion for Upgrading Sludge Handling Facilities

11.2.3.1 Use of Existing Facilities

As an upgrading technique, aerobic digestion can be carried out in existing unused tankage, such as old Imhoff tanks or old clarifiers. If unusually shaped basins are used, attention should be placed on ensuring that complete mixing will be achieved and that dead spots will be prevented. Potential dead spots can be filled and covered with concrete. Air-diffusion systems are more easily adapted to unusual basin shapes than are surface aerators.

11.2.3.2 Supplemental Aerobic Digestion

Aerobic digestion can be used in conjunction with existing anaerobic digesters. In Monroe, Wisconsin, and Corpus Christi, Texas, it has been found best to digest the primary sludge anaerobically and the waste biological sludge aerobically (33). The advantage of this segregation is that the primary sludge is not diluted by the waste biological sludge and that the anaerobically digested primary sludge subsequently filters better on a vacuum filter when it does not contain the waste biological sludge. Both digestion systems produce a stabilized nuisance-free sludge.

11.2.3.3 Conversion of Anaerobic Digesters to Aerobic Digestion

If existing anaerobic digesters are overloaded and for some reason cannot be upgraded as described in Section 11.1, they can be converted to aerobic digesters. Aerobic digesters yield similar volatile suspended solids (VSS) reductions and are relatively odor free. Hence, this conversion may be applicable to small overloaded plants in residential areas. Aerobic digestion will usually require an increase in the blower capacity of the air supply system, which would increase the yearly operating cost. However, these increased costs could be offset by savings in maintenance requirements. An alternative method of supplying the additional air is by using mechanical surface aeration where tank geometry permits.

If an existing anaerobic digester is converted to operate as an aerobic digester, the cover should be removed. Experience at one midwestern city indicated that the combination of an air diffusion system and a covered digester increased the air temperature inside the digester to 135°F and the sludge temperature to 97°F (34). At these high temperatures, objectionable odors were produced, and the addition of air caused the release of these odors to the surrounding residential area. These high temperatures, in turn, reduced the ability of the air supply to maintain the desired dissolved oxygen levels. An increase in air supply to the system was then required, which resulted in a subsequently higher liquid temperature.

11.2.4 Process Designs and Cost Estimates

Two examples are presented to illustrate upgrading of sludge handling facilities using aerobic digestion.

11.2.4.1 Example A

As a result of upgrading a trickling filter plant from 0.66 mgd to 0.88 mgd capacity, an existing two-stage anaerobic digestion system experienced operational problems. Original design data for the plant are listed in Table 11-9. Excess primary clarifier capacity was available due to an original conservative clarifier design. The excess capacity could be advantageously used as an aerobic digestion basin. The aerobic digestion of a portion of the combined sludge permitted a decreased loading to the anaerobic digestion system, resulting in a more stable operation. Modifications used for this upgrading are shown in Figure 11-7, and the upgraded design data are presented in Table 11-9.

Table 11-9

Upgrading Sludge Handling Facilities at a
Trickling Filter Plant Using Aerobic Digestion - Example A

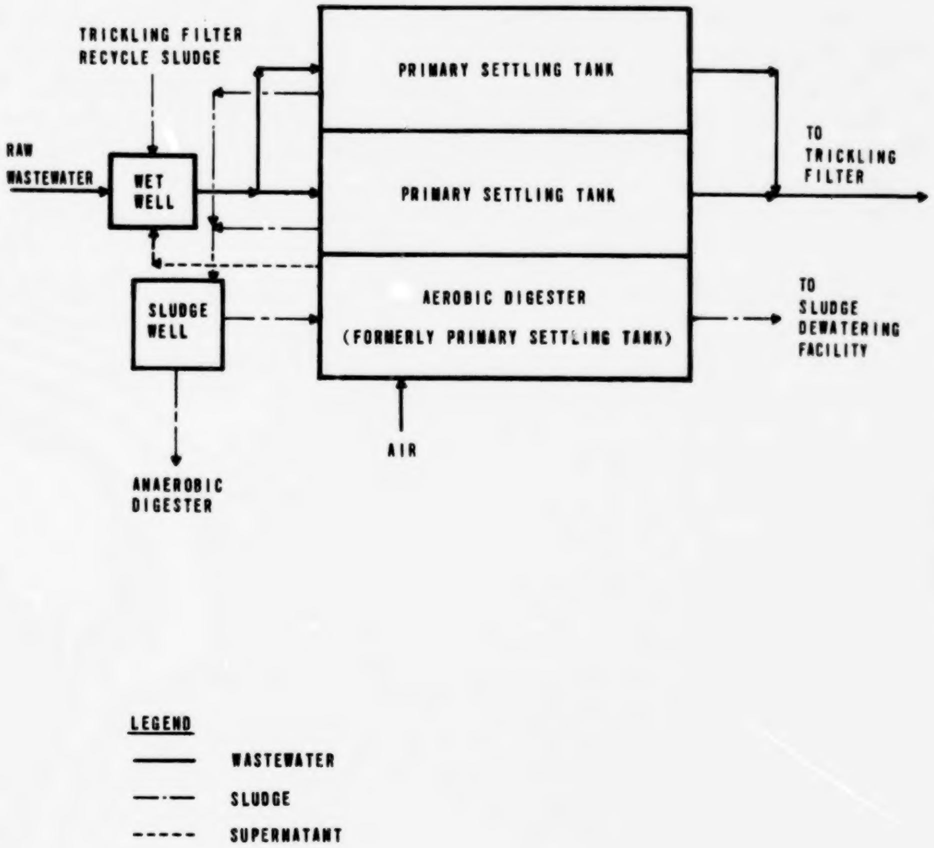
Description	Original Design	Upgraded Plant
Plant Design Capacity	0.66	0.88
Primary Clarifier (3 units)		
Overflow Rate, gpd/sq.ft.	273[1]	600[2]
Anaerobic Digester		
Detention Time, days	16.5	21.8
Volatile Solids Loading, lb. VSS/cu.ft./day	0.106[3]	0.080[3]
Sludge Solids, lbs./day	1,125	850
Sludge Volume, gpd	3,370 (4%)	2,545 (4%)
Aerobic Digester (1 unit)		
Detention Time, days	–	30
Volatile Solids Loading, lb. VSS/cu.ft./day	–	0.058[3]
Air Requirements, cfm	–	500[4]
Sludge Solids, lbs./day	–	650
Sludge Volume, gpd	–	1,950 (4%)

[1] Based on 3 units including recycle.
[2] Based on 2 units including recycle.
[3] Based on 70% VSS in sludge added to digester.
[4] 64 cfm/1,000 cu.ft.

FIGURE 11-7

UPGRADING BY USE OF SUPPLEMENTAL AEROBIC DIGESTION

EXAMPLE A


TRICKLING FILTER RECYCLE SLUDGE
RAW WASTEWATER
WET WELL
PRIMARY SETTLING TANK
PRIMARY SETTLING TANK
TO TRICKLING FILTER
SLUDGE WELL
AEROBIC DIGESTER
(FORMERLY PRIMARY SETTLING TANK)
TO SLUDGE DEWATERING FACILITY
AIR
ANAEROBIC DIGESTER
LEGEND
WASTEWATER
SLUDGE
SUPERNATANT

The capital costs for upgrading the digestion facilities were estimated at \$32,000[1] (ENR 1,500).

11.2.4.2 Example B

Due to continued operational problems with an existing anaerobic digestion facility, a community has decided to convert from anaerobic to aerobic digestion. The waste activated sludge from a 4-mgd activated sludge plant is settled along with raw wastewater in the primary clarifiers and pumped directly to the digestion system at an average solids concentration of 3 percent.

In the upgraded system, the two aerobic digesters are to be operated continuously at a total detention time of 37.4 days. This detention time is in excess of the 20 to 25 days required and will allow the aerobic digesters to handle increased solids loadings in the future. Due to the continuous operation of the digester, there will be no supernatant disposal problem associated with the digester operation. A summary of the upgraded design data is presented in Table 11-10.

Table 11-10

Aerobic Digester Upgraded Design Parameters - Example B

Description	Value
Plant Design Flow, mgd	4
Number of Digesters	2
Digester Volume (each), cu.ft.	75,000
Total Solids Added to Digesters, lbs./day	7,500
Volatile Solids Added to Digesters (70 percent), lbs./day	5,250
Sludge Volume (3 percent), gpd	30,000
Digester VSS Loading (each), lb. VSS/cu.ft./day	0.035
Hydraulic Detention Time (total), days	37.4
Aeration Requirements (each), cfm	4,500[1]

[1] (60 cfm/1,000 cu.ft.)

Both tanks are provided with diffused air equipment to supply the required amount of air. For this conversion, the digester covers were removed to ensure that the heat produced by the air diffusion system was dissipated. A simplified flow diagram is shown in Figure 11-8.

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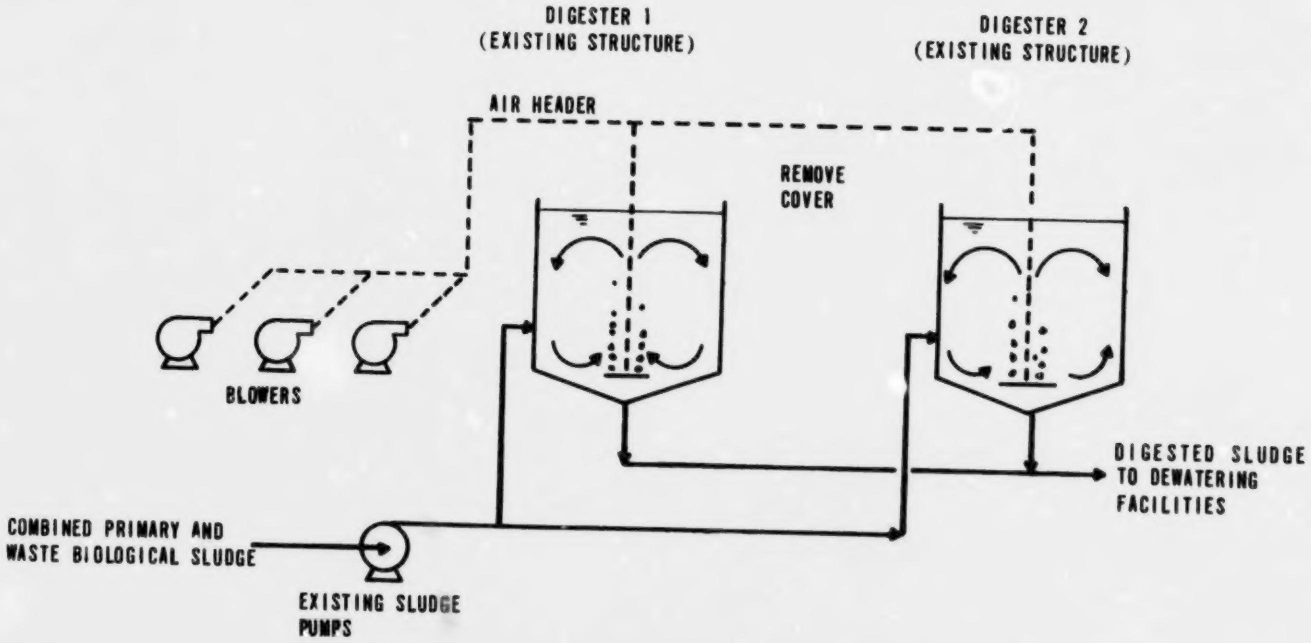
[1] These costs do not include a contingency for engineering design, bonding, and construction supervision.

FIGURE 11-8

CONVERSION OF ANAEROBIC TO AEROBIC DIGESTION

EXAMPLE B


DIGESTER 1
(EXISTING STRUCTURE)
DIGESTER 2
(EXISTING STRUCTURE)
AIR HEADER
REMOVE COVER
BLOWERS
DIGESTED SLUDGE TO DEWATERING FACILITIES
COMBINED PRIMARY AND WASTE BIOLOGICAL SLUDGE
EXISTING SLUDGE PUMPS

The estimated total capital costs for the upgrading are $175,000[1] (ENR 1500), broken down as follows:

Air System	$135,000
Renovation to Existing Tank	40,000
TOTAL	$175,000

11.3 References

1. McCarty, P.L., *Anaerobic Waste Treatment Fundamentals*. Public Works 95, No. 9, pp. 107-112 (1964).

2. Speece, R.L., and McCarty, P.L., *Nutrient Requirements and Biological Solids Accumulation in Anaerobic Digestion*. Proceedings of the International Conference on Water Pollution Research, Pergamon Press, 1962.

3. Kugelman, I.J., and Chin, K.K., *Toxicity Synergism and Antagonism in Anaerobic Waste Treatment Processes*. Presented before Division or Air, Water and Waste Chemistry, American Chemical Society, Houston, Texas, February, 1970.

4. Kugelman, I.J., and McCarty, P.L., *Cation Toxicity and Stimulation in Anaerobic Waste Treatment*. Journal Water Pollution Control Federation, 37, No. 1, pp. 97-115 (1965).

5. Lawrence, A.W., Kugelman, I.J., and McCarty, P.L., *Ion Effects in Anaerobic Digestion*. Technical Report No. 33, Department of Civil Engineering, Stanford University, March, 1964.

6. Lawrence, A.W., and McCarty, P.L., *Unified Basis for Biological Treatment Design and Operation*. Journal of the Sanitary Engineering Division, ASCE, 96, No. 3, pp. 757-778 (1970).

7. Lawrence, A.W., and McCarty, P.L., *Kinetics of Methane Fermentation in Anaerobic Treatment*. Journal Water Pollution Control Federation, 41, No. 2, Part 2, pp. R1-R17 (1969).

8. Torpey, W.N., *Loading to Failure of a Pilot High Rate Digester*. Sewage and Industrial Wastes, 27, No. 2, pp. 121-133 (1955).

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[1] These costs do not include a contingency for engineering design, bonding, or construction supervision.

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CHAPTER 12

SLUDGE DEWATERING

12.1 Vacuum Filtration

In 1967, there were slightly more than 1,500 vacuum filter installations in the United States in wastewater treatment service (1). The majority of these installations are in the larger municipal plants, where scarcity of available land often places sludge drying beds in an unfavorable economical position. The high degree of operator skill required for efficient vacuum filter operation is more likely to be available at larger plants than at smaller plants. Another factor which discourages extensive use of vacuum filters for sludge dewatering at small plants is the requirement for sludge conditioning prior to filtration. Small plants often do not have the necessary storage and handling facilities to purchase conditioning chemicals in economical bulk quantities.

12.1.1 Process Considerations

Process operating considerations for vacuum filtration include:

1. Control of feed solids concentration.
2. Chemical requirements for sludge conditioning.
3. Sludge mixing and flocculation.
4. Drum speed and drum submergence.
5. Filter fabric characteristics.

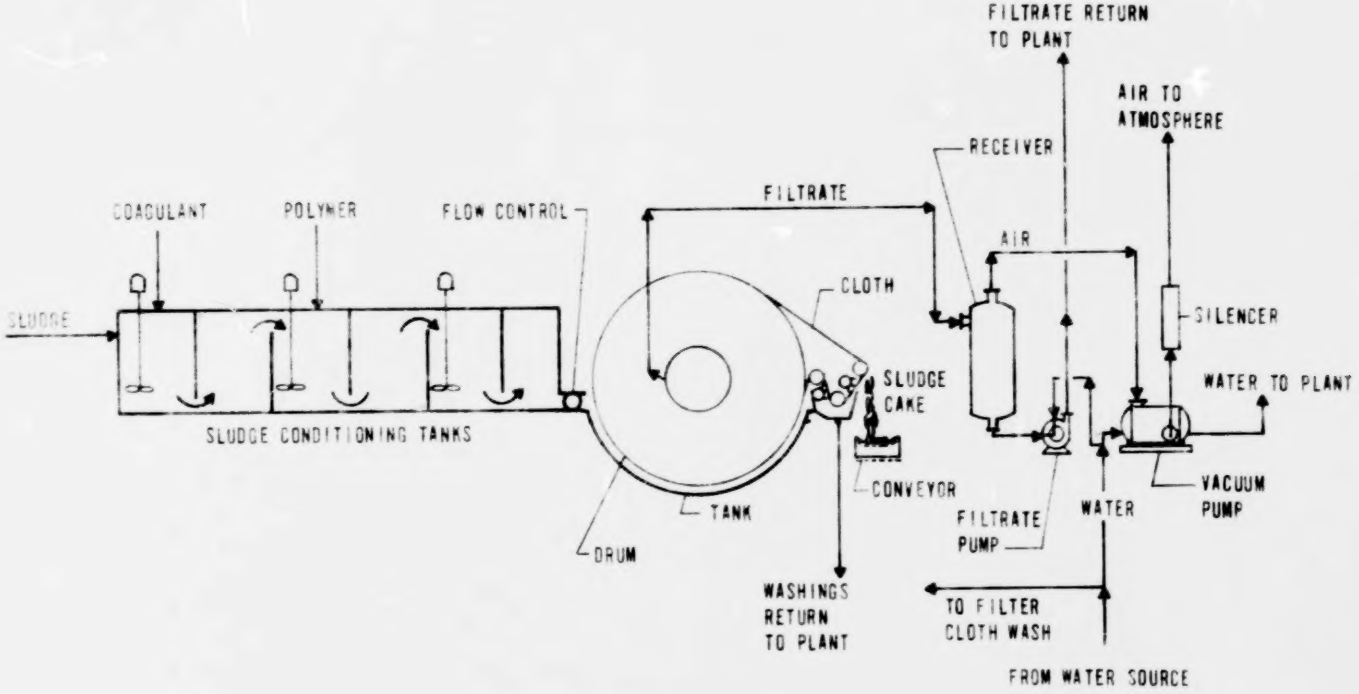
Each of these parameters affects the filter yield, economy of operation, and filter cake characteristics. The feed solids concentration can be controlled by pre-thickening the sludge. In general, the higher the feed solids concentration, the higher the filtration rate and the filter yield. The relationship is not linear, however, as was noted at one installation where doubling the feed solids concentration more than doubled the filtration rate (2).

Few, if any, raw or digested wastewater sludges can be successfully dewatered without some form of chemical conditioning using ferric chloride, lime, and/or polyelectrolytes. Proper sludge conditioning requires a determination of optimum chemical dosages. Experience and careful laboratory monitoring of the sludge characteristics are key factors in maintaining the proper chemical proportions and concentrations.

Optimum sludge mixing and flocculation under varying conditions require that sludge conditioning tanks be provided with variable-speed mixer drives, removable weirs to vary the sludge detention time, and multiple points of chemical application as indicated in Figure 12-1. The sludge slurry must be agitated sufficiently to maintain uniformity; however, excessive agitation should be avoided so that the conditioned slurry particles are not sheared and broken up.

FIGURE 12-1

TYPICAL VACUUM FILTER FLOW DIAGRAM


FILTRATE RETURN
TO PLANT
AIR TO
ATMOSPHERE
RECEIVER
FILTRATE
AIR
COAGULANT
POLYMER
FLOW CONTROL
CLOTH
SILENCER
SLUDGE
SLUDGE
CAKE
WATER TO PLANT
SLUDGE CONDITIONING TANKS
CONVEYOR
VACUUM
PUMP
TANK
FILTRATE
PUMP
WATER
DRUM
WASHINGS
RETURN
TO PLANT
TO FILTER
CLOTH WASH
FROM WATER SOURCE

Drum speed and drum submergence are important factors in the operation of vacuum filters since they affect filter yield and filter cake moisture. Increasing the drum submergence generally results in increased filter yield, but produces a filter cake with higher moisture content. Decreasing the drum speed, i.e. increasing the cycle time, has the opposite effect of decreasing the filter yield, but produces a cake with lower moisture content.

Much information is available from the various manufacturers of vacuum filters and filter media on the selection of a proper medium. The range of filter fabrics available for metal-belt or coil-type filters is limited when compared with the great number of cloths available for use with conventional drum or belt-type filters. Laboratory experimentation, using the Filter Leaf Test, provides information on filter farbic blinding, cake discharge characteristics, and filtrate quality, which is of use in selecting the best fabric for a given sludge.

12.1.2 Evaluation of the Vacuum Filtration Process

Experience has shown that there are considerable variations in filtration rate, not only between different sludge types, but also between the same types of sludges at different locations. The discrepancies in filter test results at different plants are usually related to variations in feed solids concentrations, particle size distributions, industrial waste components in the raw wastewater, etc. Obviously, then, it is of major concern to have a laboratory technique which can accurately predict the performance of a full-scale vacuum filter prior to its installation.

The Buchner Funnel Test and the Filter Leaf Test are commonly used in laboratory testing programs for estimating the filterability of sludges. When the amount of representative sludge is limited (less than 10 liters), it is advisable first to perform the Buchner Funnel Test to determine optimal chemical dosage and sludge filtration characteristics. The Filter Leaf Test can then be run at the optimum condition to determine filter yield. If a large amount of sludge is available, the Buchner Funnel Test can be eliminated and the Filter Leaf Test run instead.

The main purpose of the Buchner Funnel Test is to evaluate the optimum chemical requirements based on a comparison of the specific resistance of chemically treated sludge with that of untreated sludge. An approximate filter yield can also be calculated from the Buchner Funnel Test. Basically, the Buchner Funnel Test consists of filtering 100 ml of sludge, either raw or conditioned, through filter paper under a vacuum of 20 to 25 inches of mercury. The volume of filtrate (V, in ml) with time is noted and plotted against elapsed time/volume (t/V in sec./ml) to obtain the slope of the resulting line. Using the above information, the specific resistance of the sludge is calculated from the following equation:

$$r = 6.91 \times 10^6 \left(\frac{bA^2P}{\mu c}\right)$$

where:

r = Specific resistance, $\sec^2/\mathrm{gm}$
b = Slope of plot (V vs. t/V), $\sec/\mathrm{ml}^2$
A = Area of filter, sq.cm.
P = Filtration vacuum, psig
μ = Absolute viscosity of filtrate, centipoise
c = Initial suspended solids concentration, mg/ml

The dimensions of the variables in the above equation are in units typically measured in the laboratory. The conversion constant, 6.91×10^6, reduces the variables to units which are dimensionally consistent. The specific resistance as calculated by the above equation would be expressed as $\sec^2/\mathrm{gm}$.

The approximate filter yield can be estimated from the specific resistance data using the following relationship (3):

$$L = 0.0357 \left(\frac{100-C_f}{C_i-C_f}\right) \left(\frac{mPC_i\,(100-C_i)}{\theta R \mu}\right)^{1/2}$$

where:

L = Filter yield, lbs./sq.ft./hr.
C_i and C_f = Initial and final moisture content of the sludge, percent
m = Percentage of time for which vacuum acts during cycle
θ = Time for one drum revolution, minutes (usually between 1.5 and 5 minutes)
μ = Absolute viscosity of filtrate, centipoise
$R = r \times 10^{-7}$, $\mathrm{gm}/\sec^2$
r = Specific resistance, $\sec^2/\mathrm{gm}$
P = Filtration vacuum, psig

The Filter Leaf Test techniques are simple, and the test can be run with minimum effort. With careful laboratory techniques, results will be closely indicative of full-scale vacuum filter operation and need only be corrected by a design factor used to compensate for partial fabric blinding over a long period of operation. A scale-up design factor of 0.9 is typically used. The advantage of the Filter Leaf Test is that the filter yield is measured and not merely calculated using an empirical equation.

The Filter Leaf Test is usually performed on a 0.1 sq.ft. filter leaf. Different filter fabrics should be evaluated at a specified vacuum pressure. Varying doses and types of chemicals should also be tested to establish chemical conditioning requirements.

The main objective of the Filter Leaf Test is to evaluate the effect of different fabrics, fabric forms, and drying times on filter yield. The basic steps in performing a Filter Leaf Test consist of the following:

1. Submerge the filter leaf in the sludge slurry and apply vacuum for a fixed form time.

2. Remove the filter leaf from the sludge slurry to allow the cake to dry for a fixed drying time.

3. Remove the cake from the filter leaf and measure the weight and moisture content of the cake.

4. Measure the filtrate suspended solids to determine the filtrate characteristics.

The filter yield is then calculated using the following equation:

$$L = \frac{\text{dry sludge weight (gm) x number of filtration cycles/hr.}}{\text{453.6 x area of test filter leaf (sq.ft.)}}$$

where: L = Filter yield in lbs./sq.ft./hr.

12.1.3 Upgrading Existing Vacuum Filters

The need to upgrade an existing vacuum filter is usually due to an increase in the solids loading. Under these circumstances, existing vacuum filters are required to dewater at an increased filtration rate; otherwise, additional filtration capacity must be provided. The filtration rate of an existing vacuum filter can sometimes be increased by careful attention to operating conditions and judicious use of polyelectrolytes, where applicable, to improve sludge conditioning.

Most older filter installations were designed to use inorganic chemicals, such as ferric chloride and lime, as sludge conditioners. Within the last five years, organic polyelectrolytes have begun to replace inorganic chemicals as sludge conditioners, and these polymers offer an attractive advantage in more economical storage, handling, and feeding equipment. Polymers are also less corrosive and are frequently less expensive than inorganic chemicals within normal dosage ranges.

The yield obtained when using polymers in municipal sludge conditioning is generally higher than when using inorganic chemical conditioners. This statement is supported by studies conducted by the Dow Chemical Company (4) in which polyelectrolytes and inorganic conditioning chemicals were compared on the same sludges. The results shown in Table 12-1 indicate that polymer conditioning not only increases filter yield, but significantly reduces chemical requirements for conditioning sludge.

The use of polymers in improving the operation of vacuum filters has been practiced at treatment plants in Bay City, Michigan (5), and Kansas City, Missouri (6), among others.

The Bay City Wastewater Treatment Plant provides primary treatment for 7 mgd and produces approximately 450 tons of dry solids/year. Until 1961, raw primary sludge was

Table 12-1

Vacuum Filtration Results Comparing Inorganic Chemicals with Purifloc C-31 on Municipal Sludge

	Location	Type of Sludge	Type of Filter	Filter Media	Chemical Conditioning	Dosage lbs/ton dry solids	Solids Concentration Initial percent	Solids Concentration Final percent	Filter Yield lbs/sq ft/hr	Increase in Yield Due to Use of Polyelectrolytes percent
1	Municipal STP	Raw primary	K-S	coil	$FeCl_3$	162.4	7.57	20.1	6.91	
					Lime	166.8				
	Municipal STP	Raw primary	K-S	coil	C-31	14.0	7.0	20.0	7.53	9
2	Municipal STP	Raw primary	K-S	coil	$Fe_2(SO_4)_3$	60				
					Lime	106	9.1	24.0	8.6	
	Municipal STP	Raw primary	K-S	coil	C-31	8.4	9.6	20.0	11.5	34
3	Municipal STP	Raw primary	Eimco drum	open synthetic	$FeCl_3$	80.0				
					Lime	280.0	15	40.0	5.0	
	Municipal STP	Raw primary	Eimco drum	open synthetic	C-31	18.0	12	30.0	7.0	40
4	Municipal STP	Raw primary	Eimco drum	open synthetic	$FeCl_3$	78.0				
					Lime	390.0	11.2	39.0	3.1	
	Municipal STP	Raw primary	Eimco drum	open synthetic	C-31	20.0	10.7	34.5	3.5[1]	13
5	Atlanta, Clayton	Digested primary	D-O drum	long napped dacron	$Fe_2(SO_4)_3$	($6.06/T)	7.2	26.0	3.9	
	Atlanta, Clayton	Digested primary	D-O drum	long napped dacron	C-31	($6.58/T)	7.2	33.5	11.0	182
6	Municipal STP	Digested primary	drum	44 x 44 saran	$FeCl_3$	66.0	15.0	43.0	9.2	
					Lime	206.0				
	Municipal STP	Digested primary	drum	44 x 44 saran	C-31	17.0	15.0	32.0	25.0	172
7	Municipal STP	Elutriated/digested primary	D-O drum	napped polyester	$FeCl_3$	56.8	6.1	36.0	5.74	
	Municipal STP	Elutriated/digested primary	D-O drum	napped polyester	C-31	10.2	7.7	32.9	12.66	121
8	Municipal STP	Elutriated/digested primary	D-O drum	napped polyester	$FeCl_3$	100.0	10.4	32.7	3.88	
	Municipal STP	Elutriated/digested primary	D-O drum	napped polyester	C-31	8.0	10.1	38.6	5.94	53
9	Municipal STP	Elutriated/digested primary	Eimco drum	napped polyester	$Fe_2(SO_4)_3$	108.0	10.9	34.0	2.8	
	Municipal STP	Elutriated/digested primary	Eimco drum	napped polyester	C-31	9.0	11.1	35.0	5.5	96
10	Atlanta - South River	Elutriated/digested primary and secondary	D-O drum	long napped dacron	Lime	($8.90/T)	9.1	25.0	4.7	
					$Fe_2(SO_4)_3$	($8.79/T)	9.1	24.0	7.4	57
					C-31					
11	Municipal STP	Digested primary and secondary	Eimco	synthetic	$FeCl_3$	610.0(total)				
					Lime	or ($10.07/ton)	4.4	26.2	5.2	
	Municipal STP	Digested primary and secondary	Eimco	synthetic	C-31	18.0($6.12/ton)	4.3	23.7	5.6	8
12	Municipal STP	Elutriated/digested primary and secondary	D-O drum	napped dacron	$Fe_2(SO_4)_3$	360.0				
					Lime	120.0	8.0	28.0	5.1	
	Municipal STP	Elutriated/digested primary and secondary	D-O drum	napped dacron	C-31	22.0	9.0	25.0	7.25	42

[1] Yield intentionally kept down to avoid overloading incinerator

Source: Dow Chemical Company (4)

conditioned with ferric chloride ($FeCl_3$) and either kiln-dried pebble lime or spent carbide (calcium hydroxide formed as the result of chemical action in making acetylene). Bay City's conditioned sludge is dewatered on vacuum filters having an effective area of 150 sq.ft. The vacuum filter cake is incinerated. In 1961, polyelectrolytes were tried as sludge conditioners in an attempt to improve filter yield. Results of filter operation for 1959-1964 using $FeCl_3$, lime, and polyelectrolytes (5) are presented in Table 12-2.

These results clearly indicate that use of polyelectrolytes increased the filter yield and significantly reduced vacuum filtration operation time.

The cost of chemicals for sludge conditioning was found to be $9.93/ton of dry solids using $FeCl_3$ and kiln dried lime, $6.85/ton of dry solids when using $FeCl_3$ and carbide lime, and $7.00/ton of dry solids when using polyelectrolytes.

The following advantages were realized at Bay City when polyelectrolytes were used for sludge conditioning:

1. Equipment and floor space savings.
2. Improved housekeeping.
3. Improved safety.
4. Reduced quantities of ash, with a large reduction in ash handling and storage.
5. Reduction in operating time, with resulting savings in operating and maintenance costs.

In Kansas City, 115 mgd of wastewater are treated in two plants, and the resulting sludge is pumped to a central location for sludge dewatering (6). Polyelectrolytes were selected for sludge conditioning in the dewatering operation. Dewatered sludge is incinerated. The specifications for sludge dewatering included: a filter capacity of 2,300 lbs./hr. for each of eight filters, a filtration rate of 6.2 lbs./sq.ft./hr. at minimum solids concentration, and a maximum moisture content of 75 percent in the cake. The cost of polymer conditioning was initially estimated at $4.30/ton of dry solids. During the first six months of operation, however, the actual cost of polymer was $10.61/ton dry solids. This cost was reduced by $2.20/ton of dry solids during the subsequent four-year period of operation.

Despite the higher than estimated cost for sludge conditioning, the experiences of the Kansas City Treatment Plant with polyelectrolytes have been satisfactory, since both the filter yield and cake moisture content specifications are routinely met.

One area where polyelectrolytes have not been as effective as inorganic chemical conditioning is in the vacuum filtration of waste activated sludge directly from the underflow of a secondary clarifier (without thickening). Preliminary information from an oxygen aeration study conducted at Batavia, New York, indicates that the optimum conditioning was ferric chloride at a dosage of 200 lbs./ton of dry solids (7). The waste activated sludge concentration from the oxygen aeration system varied between

Table 12-2

A Comparison Between Lime/Ferric Chloride and Polyelectrolytes for Conditioning Raw Primary Sludge

Year	Dry Solids	Solids in Feed Sludge	Filter Yield	Amount of Substance Added pounds						Solids in Cake	Solids Recovery	Operation Time
				Lime	FeCl3	C-31	C-32	C-149	A-21			
	tons	percent	lbs. sq.ft. hr.							percent	percent	hours
1959	461	11.2	3.1	162,000	31,000	—	—	—	—	40.1	64.1	2,125
1960	580	11.2	3.1	225,000	44,000	—	—	—	—	39.0	62.1	2,420
1961-62	424	10.9	5.3	—	—	—	5,562	—	—	35.9	75.6	1,119
1962-63	415	10.7	5.5	—	—	10,300	—	—	—	34.5	73.7	1,114
1963-64	437	10.2	6.3	—	—	—	Total of all three polymers 7,671			34.6	75.9	1,301

Source Sherbick (5).

2 and 3 percent. Filter yield increased from 1 to 5 lbs./sq.ft./hr. as the cycle time was decreased from 6 to 2 minutes/revolution. The moisture content varied between 75 and 85 percent, with the higher moisture content generally corresponding with lower cycle times. Likewise, in the City of Milwaukee, unthickened (diffused air) waste activated sludge is vacuum filtered using ferric chloride (8). The filter yield ranges between 1 and 3 lbs./sq.ft./hr. for an average cycle time of 3.5 to 4.0 minutes/revolution (cake moisture content 80 to 85 percent).

12.1.4 Process Designs and Cost Estimates

The following example will serve to illustrate design and cost considerations when upgrading vacuum filter installations by converting from inorganic chemical to polyelectrolyte sludge conditioning.

An existing vacuum filter installation annually conditions and filters 300 tons (dry basis) of mixed digested primary and secondary sludge with a filter yield of 5 lbs./sq.ft./hr. In the past, sludge conditioning has been accomplished using 65 lbs. of ferric chloride and 200 lbs. of lime per ton of dry solids. To reduce the costs involved in bulk chemical handling and to increase filter yield, the applicability of a polyelectrolyte system for sludge conditioning is investigated.

The optimum polyelectrolyte dosage is found to be 20 lbs./ton of dry solids, added to the digested sludge in a 1-percent solution. The polyelectrolyte addition results in a subsequent filter yield of 8 lbs./sq.ft./hr. This upgrading procedure enables the vacuum filter to decrease its operating time by over 60 percent, thereby decreasing operational and maintenance costs.

The capital cost for the polyelectrolyte application system is estimated at $6,000 (ENR Index 1500). This cost includes all required tankage, pumps, and mixers. In actuality, this cost is probably high since some of the equipment could probably be salvaged from the existing inorganic chemical addition system.

12.2 Drying Beds

The dewatering of digested sludge on drying beds has long been practiced in the United States. Historically, sludge drying beds have been used for communities of many sizes, as illustrated in the following table (9):

Table 12-3

Distribution of Sludge-Drying Beds
by Population Size Groups

Population Size Group	Total Number of Plants	Number with Sludge-Drying Beds	Percentage with Drying Beds
Less than 1,000	3,780	1,237	33
1,000 to 5,000	4,990	2,759	55
5,000 to 10,000	1,437	857	59
10,000 to 25,000	1,164	659	57
25,000 to 50,000	473	217	46
50,000 to 100,000	269	121	45
More than 100,000	452	196	43
TOTAL	12,565	6,046	48

The popularity of drying beds for dewatering sludge is due to their operational flexibility, simplicity, and low maintenance costs. Disadvantages include their large land requirement and inability to dewater effectively during inclement weather.

12.2.1 Process Considerations

One of the difficulties in developing a rational design for sludge beds is the multitide of variables which affect the drying rate of sludges when applied to sand beds. In practice, it is difficult to isolate these variables and evaluate them quantitatively. Some of the more important variables are (10):

1. Climate and atmospheric conditions.
   a. Temperature
   b. Humidity
   c. Rainfall
   d. Wind velocity
   e. Barometric Pressure
   f. Solar Radiation
2. Depth of sludge application.
3. Presence or absence of coagulants.
4. Sludge moisture content.

5. Source and type of sludge.
6. Extent of sludge digestion.
7. Sludge age.
8. Sludge composition.
9. Sludge concentration when applied.
10. Sludge bed construction.

Notwithstanding the magnitude of the problems involved, some generalizations concerning the applicability of these factors can be made. When possible, decisions regarding the specific effect of any or all of these factors should be based on bench-scale testing.

Quon and Johnson (11) have indicated that well-digested sludge should be applied to drying beds in depths of 6 to 9 inches, with 8 inches appearing to give optimum drying rates.

Sludge should be properly digested before being applied to the drying beds. Raw or poorly digested sludge dewaters slowly on the drying beds and produces strong odors. Sludge that has been overly digested exhibits high density which also impairs drainage. Aerobically digested sludge usually has good dewatering characteristics and, when applied to sand drying beds, drains well (2).

It has been widely accepted that under normal conditions, practically all of the drainage of digested sludge occurs during the first three days following the filling of the drying bed (1). After this initial period, it was felt that evaporation was largely responsible for additional dewatering of the sludge. Recent studies indicate this is not the case. In an extended study, it was found that the initial rate of drainage was small, but that it increased with time (11). After approximately three days, the drainage rate increased and the sludge surface dropped substantially. This phenomenon is explained by considering that air trapped in the voids of the sand bed is not free to move and thus impedes the initial flow of water through the filter. Eventually, this air is liberated, allowing a greater flow to pass through the sand bed. After a period of maximum drainage, the drainage rate gradually decreases due to the build-up of solids on the sand surface, which offers resistance to further filtration. Once this point is reached, evaporation from the free water surface accounts for further dewatering. Experiments in some installations have shown that tile-drained sludge beds dry 25 percent faster than beds with an impervious bottom (11).

In certain areas with adverse climatic conditions, the use of glass-covered beds, while expensive, has been found to increase the total output of dewatered sludge by 100 percent (12). In many cases, this increase makes glass-covered beds considerably cheaper in the long run. Adequate ventilation must be provided in constructing covered beds so that maximum evaporation rates may be maintained.

12.2.2 Design Basis

Present-day design practices are still based largely on comparisons with existing plants in the area, or upon empirical recommendations of various regulatory agencies. The following sludge drying bed area requirements are specified in the Ten-States Standards for domestic wastewater treatment plants located in northern United States (13).

Table 12-4

Sludge-Drying Bed Area Requirements

Type of Sludge	Area of Drying Beds sq.ft./capita Open Beds	Area of Drying Beds sq.ft./capita Covered Beds
Primary digested	1.0 to 1.5	0.75 to 1.0
Primary and humus digested	1.25 to 1.75	1.0 to 1.25
Primary and activated digested	1.75 to 2.5	1.25 to 1.5
Primary and chemically precipitated digested	2.0 to 2.5	1.25 to 1.5

In the southern United States, reduced areas are often practical because of more favorable climatic conditions.

12.2.3 Upgrading Existing Facilities

It is possible to upgrade an overloaded sludge drying bed by the following methods:

1. Improving the performance of upstream facilities, e.g., thickeners, digesters.
2. Adding chemicals to increase sludge dewatering.
3. Covering open beds wherever climatic conditions adversely affect performance.

Chemicals such as alum, ferric chloride, and, more recently, polyelectrolytes have been used as flocculants to improve dewatering capacity of sludge drying beds. The use of these chemicals increases the permissible annual sludge loadings to the drying beds by increasing the number of sludge draws per year.

In general, the chemicals allow greater amounts of water to drain from the sludge, thereby decreasing the amount of water to be removed through the slower evaporation process. Bed loadings for chemically treated and untreated sludge should be evaluated by laboratory and field testing to determine the effectiveness of chemical addition on sludge dewatering. Buchner Funnel Tests can predict dewatering rates on drying beds in the same manner as that predicted for vacuum filter performance (14). Care must be taken to avoid adding excess amounts of chemicals, which might bind sand particles and lower dewatering rates.

The available coagulants are not equally effective for sludge dewatering. Alum has been used successfully at a dosage of 1 lb. of alum per 100 gallons of digested sludge (1). On the other hand, polyelectrolyte has been used at dosages as low as 0.05 lb. per 100 gallons of digested sludge (10).

In many parts of the country, the common practice is to cover sludge beds to protect them from rainfall and severe winter conditions. Recent work in northern Texas indicated that, during the dry season, covers retarded the drying rate rather than accelerating it (10). These data point to the fact that drying of sludge under covered conditions is not necessarily advantageous when weather conditions are more favorable for natural drying.

12.2.4 Process Designs and Cost Estimates

An open drying bed at an existing activated sludge plant was originally designed based on a population equivalent of 20,000 and a land requirement of 2 sq.ft./capita, and was loaded at a rate of 10 lbs. of dry solids/yr./sq.ft. As a result of upgrading secondary treatment units, it was necessary to increase the loading to 15 lbs./yr./sq.ft. to accommodate increased sludge quantities. Two alternatives were available for upgrading the existing drying beds. It was possible either to cover the beds and reduce the area requirements to approximately 1.35 sq.ft./capita, or to add 1 lb. of alum per 100 gallons of digested sludge to decrease the drying time by approximately 50 percent.

Covering the drying beds was estimated to cost $200,000, or $2,000/ton dry solids/yr. of increased cake yield (ENR Index 1500). The alum slurry feed system and flocculation tank was estimated at $28,000, or $280/ton dry solids/yr. of increased cake yield.

Based on comparison of these capital cost estimates for upgrading sludge drying beds, it would appear that chemical addition would be the most economic alternative provided that the climatic conditions would not adversely affect its operation. In order to make a definite conclusion, it would be necessary to compare yearly operating costs which would include chemical costs.

12.3 Centrifugation

Centrifuges have been used for many years by various industries for clarifying liquids, concentrating solids, separating immiscible liquids, and purifying oils. However, their use in the wastewater field for sludge dewatering is not as widespread as is the use of vacuum filters. Recent improvements in centrifuge design, efforts by the centrifuge industry to enter the wastewater treatment field, and broader dissemination of centrifuge performance data have encouraged increased use of centrifuges for thickening and dewatering of primary, secondary, and combined wastewater sludges. Centrifuges have good potential for upgrading overloaded solids handling facilities due to their flexibility in operation and lesser space requirements compared to vacuum filters.

12.3.1 Types of Centrifuges

There are three general classifications of centrifuges that can be applied to sludge thickening and dewatering: solid-bowl, disc, and basket centrifuges. These are illustrated in Figure 12-2 (15). The capabilities of these units in processing wastewater sludges are summarized in Table 12-5 (16).

Table 12-5

Summary of Centrifuge Characteristics

Description	Centrifuge Description		
	Solid-Bowl Continuous	Basket Batch	Disc Continuous
Bowl diameter, in.	6 to 60	12 to 60	8 to 30
Flow rate, gpm	1 to 200	100 max.	10 to 300
Solids in feed, percent	*	0.1 to 30	0.1 to 10
Solids Discharged	1 to 15 tons/hr.	1,000 lbs.-max.	10 to 3,000 gal./hr.
Speed, rpm	1,000 to 6,000	2,500 max.	4,500 to 10,000
Centrifugal Force, G	3,200 max.	2,000 max.	12,000 max.
Motor horsepower	5 to 250	100 max.	10 to 125

*Any liquid or slurry which can be pumped.

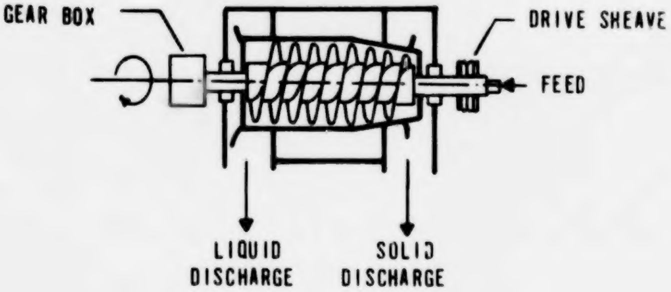
Source: Townsend (16).

The most popular type of centrifuge today is the solid-bowl because of its dependable performance and low maintenance requirements. The solid-bowl machine has a spinning cylinder which causes particles to settle out along its inner wall. Solid-bowl centrifuges are especially suited to dewatering primary wastewater sludge and mixtures of primary and waste biological sludge. They are also able to dewater waste biological sludge alone, but some form of polymer addition is required in order to operate at an economical feed rate and to obtain solids concentrations above 5 or 6 percent. For most sludges, to achieve solids recovery in the range of 80 to 95 percent with a solid-bowl centrifuge requires the addition of polymers to the sludge.

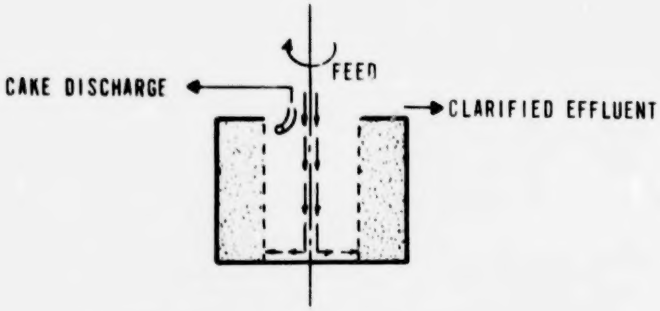
The basket centrifuge is a tubular type centrifuge with a solid bowl and, therefore, is similar to the solid-bowl centrifuge in that the solids settle out along the inner wall due to centrifugal force. The solids are removed on an automated batch basis. Because of the large bowl diameter, the basket centrifuge is operated at slower speeds. The centrifuge can be operated on automatic cycle for programmed filling and skimming. Abrasion occurs only with the skimming tool. Hence, for the most part, this is a low-speed, low-maintenance unit.

The application of basket centrifuges to the wastewater treatment field is relatively new. Field tests of this unit have been successful in thickening waste activated sludge and indicate

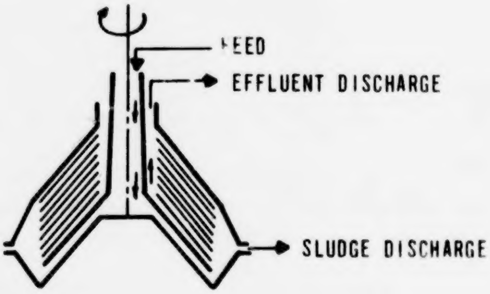
FIGURE 12-2
VARIOUS CLASSIFICATIONS OF CENTRIFUGES (15)


GEAR BOX
DRIVE SHEAVE
FEED
LIQUID DISCHARGE
SOLID DISCHARGE

SOLID BOWL CENTRIFUGE


FEED
CAKE DISCHARGE
CLARIFIED EFFLUENT

SOLID BOWL BASKET CENTRIFUGE


FEED
EFFLUENT DISCHARGE
SLUDGE DISCHARGE

DISC TYPE CENTRIFUGE

that this unit may be increasingly utilized in small plants to improve sludge dewatering operations. Concentrations of 9 to 10 percent solids can be produced, without the use of polymers, with solids recoveries of 80 to 90 percent.

The disc centrifuge utilizes a vertical disc stack, with subsequent sludge discharge through nozzles located around the periphery of the disc stack. The use of disc centrifuges for thickening in Sioux Falls, S.D. (17) and Chicago, Ill. (18) resulted in plugging of the disc stack and nozzles. It has been found that these problems can be minimized by using screens in the feed line to the centrifuge. Recently, it has been reported that disc centrifuges can increase the solids concentrations of waste activated sludge from 0.5-1.0 percent to a concentration of 5.0-6.0 percent without the use of polymers (17).

12.3.2 Process Considerations

Process variables for centrifugation are feed rate, sludge solids characteristics, feed consistency, temperature, and chemical additives. Machine variables are bowl design, bowl speed, pool volume, and conveyor speed (2). Major factors of importance in the product sludge are cake dryness and solids recovery. To increase cake dryness, the following guidelines are important (2):

1. Increase feed rate.
2. Decrease feed solids concentration.
3. Increase temperature.
4. Do not use flocculants.
5. Increase bowl speed.
6. Decrease pool volume.
7. Decrease conveyor speed.

Guidelines for increasing solids recovery are as follows (2):

1. Decrease feed rate.
2. Increase feed solids concentration.
3. Increase temperature.
4. Use flocculants.
5. Increase bowl speed.
6. Increase pool volume.
7. Decrease conveyor speed.

The above–mentioned guidelines indicate that most of the variables which improve cake dryness tend to decrease the solids recovery. This is an important feature of centrifuge operation. Therefore, operation of a centrifuge should be optimized to obtain the desired balance between cake dryness and solids recovery.

The following advantages are associated with the use of a centrifuge:

1. Capital cost is low in comparison with other mechanical equipment.
2. Operating costs are moderate, provided flocculants are not required.
3. The unit is totally enclosed so that odors are minimized.
4. The unit is simple and will fit in a small space.
5. Chemical conditioning of the sludge is often not required.
6. The unit is flexible in that it can handle a wide variety of solids and can function as a thickening as well as a dewatering device.
7. Little supervision is required.
8. The centrifuge can dewater some industrial sludges that cannot be handled by vacuum filters.

Disadvantages of its use are:

1. Without the use of chemicals, solids capture is often poor.
2. Chemical costs can be substantial.
3. Trash must often be removed from the centrifuge feed by screening.
4. Percent cake solids are often lower than those resulting from vacuum filtration.
5. Maintenance costs are high.
6. Fine solids which escape the centrifuge (in the centrate) may resist settling when recycled to the head of the treatment plant and gradually build up in concentration and eventually raise effluent solids levels.

12.3.3 Design Considerations

Centrifuges are usually selected on the basis of tests with smaller, geometrically-similar machines. The use of a continuous centrifuge for testing purposes is not realistic if sludge supply is limited, as when evaluating a small pilot plant. In the past, no laboratory procedure has been available to predict prototype performance. This is in direct contrast to the laboratory procedures, namely the Buchner Funnel Test and the Filter Leaf Test, which have been developed to assess the dewatering characteristics of the vacuum filter.

Recently, Vesilind (19) developed a laboratory procedure which makes it possible to predict prototype centrifuge performance on the basis of percent solids recovery. By trial and error, it was found that the following model relates laboratory to prototype data:

$$\text{Estimated Percent Recovery} = 100 \left(\frac{C_f - S}{C_f}\right)\left(\frac{P}{100}\right)^{0.1}$$

where:

C_f = Feed solids concentrations, mg/l
S = Centrate solids concentration, mg/l, measured after spinning in a laboratory centrifuge at a desired centrifugal force and appropriate time.
P = Percent of sludge not penetrated as determined from a sludge penetrometer.

No attempt was made to determine percent cake solids in this study. However, it is known that percent solids recovery and cake solids have an inverse relationship, i.e., the higher the solids recovery, the lower the percent cake solids. The numerical difference between predicted and actual solids recovery from Vesilind's formula can be expected to be within plus or minus 10 percent.

12.3.4 Centrifuge Performance in Sludge Thickening and Dewatering

Table 12-6 contains operating data supplied by various centrifuge manufacturers as well as that reported in the literature for various combinations of municipal wastewater sludges. These data indicate that the solid-bowl centrifuge is the most adaptable to the various combinations of wastewater sludges.

Raw primary and digested primary sludges dewater easily. With polymer addition, a centrifuge can produce 25 to 40 percent cake solids with better than 90 percent recovery. When trickling filter sludge is added to either of these sludges, the percent cake solids drops to 20 to 25 percent, and the polymer dosage to obtain 90 percent recovery increases. Factors responsible for this loss in efficiency include lower feed solids and the less favorable dewatering characteristics of biological sludge compared to primary sludge.

Waste activated sludge, by itself without conditioning, is difficult to thicken or dewater. This is readily evident in Table 12-6 and is not peculiar to the centrifuge process. A disc centrifuge can thicken waste activated sludge to 5 to 7 percent recovery without polymers. A basket centrifuge can also thicken waste activated sludge, but large amounts of polymer are required to make the sludge scrollable. To obtain 8 to 10 percent cake solids, polymers at a cost of $15 to $20/ton of dry solids are required to obtain greater than 90 percent recovery.

The use of centrifuges for sludge dewatering has been considered recently by several municipalities. At one large southeastern city, a centrifuge test program was conducted to determine the applicability of centrifuge dewatering of raw primary, digested primary, co-settled raw primary and waste activated, mixed digested, and primary digested plus thickened waste activated sludges (20). With a solid-bowl centrifuge, 55 to 85 percent recovery was obtained without the use of polyelectrolytes depending upon the feed rate. Recovery levels of 85 percent or better were achieved with 0.5 to 5 pounds of strong cationic polyelectrolyte per ton of dry solids for combined raw primary and waste activated sludges, primary digested plus waste activated sludges, and mixed digested sludges. Thickening of waste activated sludge using a disc machine was found to be feasible and produced 5 to 7 percent solids without polyelectrolytes. Problems were encountered in trying to use a solid-bowl centrifuge on the combination of thickened waste activated and primary digested sludges. Higher levels of polyelectrolyte were required as the proportion of activated sludge to primary sludge increased. When operating with a mixture of 2/3 primary digested sludge and 1/3 thickened waste activated sludge on a dry solids basis, 50 to 60 percent recovery was achieved without polyelectrolytes. To increase recovery to 80 to 90 percent, polyelectrolyte dosages of 10 to 20 pounds/ton of dry solids were required.

Table 12-6

Centrifuge Performance Data

Type of Sludge	Centrifuge Type	Capacity gpm	Percent Feed Solids	Percent Cake Solids	Percent Recovery	Polymer Requirement		Reference
						$/ton	lbs./ton	
Primary - Raw	Solid Bowl			30-40	70-80		None	20
Primary - Raw	Solid Bowl			30-40	95		5-10	20
Primary - Raw	Solid Bowl			25-35	90+	2-4		20
Primary - Raw	Solid Bowl			28-40	70-90	None		20
Primary - Raw	Solid Bowl			30-40	50-90		1.5-2.5	20
Primary - Raw	Solid Bowl	23-26	9-12	28-50	65-80	None		20
Primary - Raw	Solid Bowl	23-36	9-12	28-50	80-98		1.0-2.5	20
Primary - Raw	Solid Bowl			32-37	84-93	6.4		21
Primary - Raw	Solid Bowl			25-28	88-95	3-5		21
Primary - Raw	Solid Bowl			26-37	85-96	2-7		21
Primary - Raw	Solid Bowl	27.5	8.8		44	None		23
Primary - Raw	Solid Bowl	27.5	8.4		84	3.94		23
Primary - Raw	Solid Bowl	27.5	9.2		97	5.72		23
Primary - Digested	Solid Bowl			30-40	70-90		None	22
Primary - Digested	Solid Bowl			30-40	95		5-10	22
Primary - Digested	Solid Bowl			24-30	90+	3-6		22
Primary - Digested	Solid Bowl			25-35	70-85	None		22
Primary - Digested	Solid Bowl			20-30	85-90		3-6	20
Primary - Digested	Solid Bowl		6	22-28	75-85		0	20
Primary - Digested	Solid Bowl		6	22-28	85		0.5-1	20
Primary - Digested	Solid Bowl			26-30	81-90	8		21
Primary - Digested	Solid Bowl			22-30	87-95	3-7		
Raw Primary + T.F.	Solid Bowl			24-30	90+	3-6		22
Raw Primary + T.F.	Solid Bowl			25-35	65-75	–		22
Raw Primary + T.F.	Solid Bowl	9.9-22	9.6	23	96-100	4-9		21
Raw Primary + T.F.	Solid Bowl			20-26	82-96	5-8		21
Raw Primary + T.F.	Solid Bowl			21-25	83-90	4-10		21
Raw Primary + T.F.	Solid Bowl			20-25	85-94	6-9		21
Digested Primary + T.F.	Solid Bowl		5.5	25+	40	None		23
Digested Primary + T.F.	Solid Bowl		5.6	20.2	90	6.74		23
Digested Primary + T.F.	Solid Bowl			20-28	90+	4-8		22
Digested Primary + T.F.	Solid Bowl			22-36	60-75	–		22
Digested Primary + T.F.	Solid Bowl			25	85		2.5-3.5	20
Digested Primary + T.F.	Solid Bowl			18-25	60-85	None		21
Digested Primary + T.F.	Solid Bowl			18-25	95+	8-16		21
Raw Primary + Waste Activated	Solid Bowl	40-80		18-20	40-60		None	20
Raw Primary + Waste Activated	Solid Bowl	40-80		18-20	85		4-5	20
Waste Activated (after roughing filter)	Disc	50-80	0.7	5-7	93-87	None		20
Waste Activated (after roughing filter)	Disc	60-270	0.7	6.1	97-80	None		20

Table 12-6
(continued)

Type of Sludge	Centrifuge Type	Capacity gpm	Percent Feed Solids	Percent Cake Solids	Percent Recovery	Polymer Requirement		Reference
						$/ton	lbs./ton	
Waste Activated	Disc	150	0.75-1.0	5-5.5	90+	None		22
Waste Activated	Disc	400		4.0	80	None		22
Waste Activated	Basket	33-70	0.7	9-10	70-90	None		20
Waste Activated	Solid Bowl	10-12	1.5	9-13	90			24
Conventional	Solid Bowl			8-10	90-100	15-20		22
Thickened	Solid Bowl			5-15	90-100	5-10		22
Contact Stabilization	Solid Bowl			5-12	90+	2-5		22
or Extended Aeration	Solid Bowl			5-12	70-85	None		22
Thickened Waste Activated Sludge (by Disc Centrifuge)	Solid Bowl	40-60	5-6	15-17	50-65		6-7	20
	Solid Bowl	30-80	5-6	15-17	85		11-17	20
Raw Primary + Waste Activated	Solid Bowl			15-23	90+	5-10		22
	Solid Bowl			19-25	50-75	None		22
	Solid Bowl			18-24	50-80	None		25
	Solid Bowl			18-24	95	6-20		25
Digested Primary	Solid Bowl	70-140	4	>17	20-50		None	20
	Solid Bowl		4	>17	85		3-8	20
Digested Primary + Waste Activated	Solid Bowl	50-120		22-30	80-95		2.4-10	20
	Solid Bowl	30-150		25-32	50-75	None		20
	Solid Bowl		2.8	15-19	85-90	12		22
	Solid Bowl			15-20	50-60	None		22
	Solid Bowl			15-20	85-100	10-35		22
	Solid Bowl			18-24	50-70	None		25
	Solid Bowl			18-24	95	10-20		25
Digested Primary + Thickened Waste Activated	Solid Bowl	40-140	6	20-33	40-80	None		20
	Solid Bowl	30-80	6	20-22	85	13-17		20
Aerobic Digested (contact stabilization)	Solid Bowl			10-14	90+	5-10		22
	Solid Bowl			10-14	50-60	None		22
Heat Treatment Sludge								
Zimpro	Solid Bowl			30-45	85-90	None		22
Porteous	Solid Bowl			30-50	85-90	None		22
Chemical Sludges								
Lime-Phosphate	Solid Bowl			35-40	75-85	None		22
Lime-Treatment in Primaries	Solid Bowl			40-45	96-98		1.0	20
Tertiary Phosphate Removal (Chemical + Waste Activated)	Solid Bowl			16-20	85-90		3.4	22

At a southwestern community, field tests were conducted to determine the applicability of centrifugation for dewatering combined primary and secondary digested sludges (20). The results indicated that a solid-bowl centrifuge could be used to replace concrete drying beds. With the use of a strong cationic polyelectrolyte at a concentration of 3 to 4 lbs./ton of dry solids, the sludge could be dewatered to 17 to 18 percent with a solids recovery of 85 percent. Solids recovery was increased to 98 to 99 percent recovery when the polyelectrolyte dosage was raised to 5 to 6 lbs./ton of dry solids.

In El Paso, Texas, a land problem necessitated replacement of the sludge drying beds (26). Centrifuges were used to dewater the digested sludge at a 6 to 7 percent feed solids concentration. The centrifuges produced 20 to 22 percent cake solids with 85 to 90 percent recovery, at a polyelectrolyte dosage of 2 to 3 lbs./ton of dry solids.

The use of a cationic polyelectrolyte was evaluated for a combination of 80 percent raw Imhoff and 20 percent digested sludges (20). The cationic polyelectrolyte was effective in improving sludge dewatering at economical dosage levels (2 to 3 lbs. / ton of dry solids). With an average feed solids concentration of 8 percent, a cake solids of 35 percent at 95 percent solids recovery was obtained.

Centrifuges can also be used to replace vacuum filters for sludge dewatering. Large plants are particularly interested in evaluating the centrifuge as an alternative to other mechanical sludge dewatering devices. Savings attributed to decreased operational and maintenance costs have been noted. Pre-treatment devices to further decrease maintenance costs are also evaluated. A cyclone can be used ahead of centrifuges to remove a large fraction of the sand and other abrasives. In addition, screening of the centrifuge feed material and/or use of solids grinding is recommended to eliminate the possibility of conveyor feed zone pluggage.

A disc centrifuge has been field tested for thickening waste activated sludge at an eastern Pennsylvania community with good success (27). Using a 30-in. centrifuge with a 150-hp motor and 300-gpm feed rate, the disc centrifuge produced a 5 percent underflow with 90 percent solids recovery when the Sludge Volume Index was generally less than 100. Since the plant did not have primary treatment, it was necessary to install a screening device ahead of the centrifuge. The screening effectiveness was demonstrated in that the nozzles of the centrifuge did not plug. The combination of effective screening and a patented recirculating system (providing a larger nozzle size) was instrumental to the good performance.

12.3.5 Use of Centrifuge for Upgrading Sludge Handling Facilities

The performance of centrifuges in various applications clearly indicates that centrifugation should be considered when the upgrading of solids handling facilities is required. Centrifuges are a flexible upgrading device because of their applicability in both the thickening and the dewatering of various mixtures of sludges.

When used as a thickening device, a centrifuge can upgrade an overloaded anaerobic digester by reducing the volume of feed sludge, thereby increasing digester detention time. In addition, centrifuges can also be used to supplement existing overloaded gravity thickeners. When used as a dewatering device, they can supplement existing overloaded vacuum filters.

12.3.6 Process Designs and Cost Estimates

Two examples are given to illustrate the upgrading of thickening and dewatering facilities through the use of centrifuges.

12.3.6.1 EXAMPLE A

Anaerobic digestion facilities at an existing activated sludge plant are overloaded; as a result, the detention time in the digesters has been reduced to 11.25 days, thereby causing unstable operation of the digesters.

To upgrade the digestion facilities, it is decided to thicken the waste activated sludge by using a disc centrifuge, as shown in Figure 12-3. It is assumed that polyelectrolyte addition is not necessary. The volume of the waste activated sludge, 117,000 gpd at 1 percent solids (9,800 lbs. dry solids/day) before centrifugation, is reduced to 29,250 gpd at 4 percent solids by the centrifugation step. Combination of the thickened waste activated sludge with 36,800 gpd of 5 percent primary sludge results in an overall increase in digester detention time to approximately 17 days.

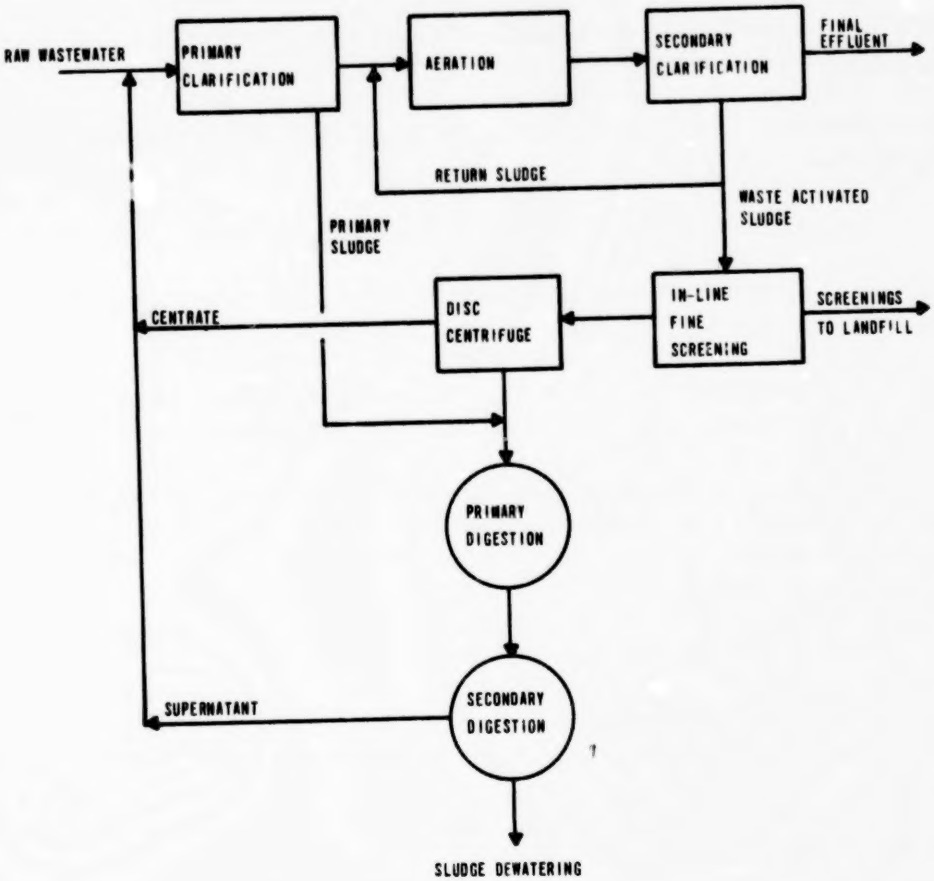
The capital cost for this upgrading procedure is estimated at $215,000 (ENR 1500) ($17,200/ton of total plant dry solids/day, or $43,900/ton of waste activated dry solids/day). This cost includes one standby disc centrifuge, in-line screens, sludge pumps, and appurtenances, but does not include an allowance for engineering design, bonding, and construction supervision.

12.3.6.2 EXAMPLE B

Vacuum filter facilities at an existing activated sludge plant are overloaded due to a recent increase in plant capacity from 10 to 20 mgd. Space limitation at the plant prohibits the installation of another filter.

It is decided to add a centrifugation facility to supplement existing vacuum filter facilities. Fifty percent of the mixed digested sludge, 12,000 lbs./day of dry solids (20 gpm by volume), is to be treated in two alternately used solid-bowl centrifuges. At an estimated 5 percent feed solids and with a polymer dosage of 3 to 6 lbs./ton of dry solids, it is expected that the solid-bowl centrifuge will produce 20 to 25 percent cake solids at 85 percent solids recovery. Facilities for sludge dewatering include 2 solid-bowl centrifuges, each capable of handling 15 to 20 gpm, sludge feed pumps, polymer addition facilities, and other necessary appurtenances.

FIGURE 12-3
EXAMPLE A
UPGRADING DIGESTION BY THICKENING WITH DISC CENTRIFUGE


RAW WASTEWATER
PRIMARY CLARIFICATION
AERATION
SECONDARY CLARIFICATION
FINAL EFFLUENT
RETURN SLUDGE
WASTE ACTIVATED SLUDGE
PRIMARY SLUDGE
CENTRATE
DISC CENTRIFUGE
IN-LINE FINE SCREENING
SCREENINGS TO LANDFILL
PRIMARY DIGESTION
SUPERNATANT
SECONDARY DIGESTION
SLUDGE DEWATERING

The capital cost for this upgrading procedure is estimated at $218,000 (ENR 1500) ($18,175/ton of total plant dry solids/day, or $36,350/ton of dry solids/day actually processed by the centrifuges). This cost does not include an allowance for engineering design, bonding, and construction supervision.

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CHAPTER 13

CASE HISTORIES OF TREATMENT PLANT UPGRADING

13.1 General

The capabilities and limitations of various unit processes have been discussed in preceding chapters. However, in any wastewater treatment plant, the operation of an individual process affects the operation of other processes in the treatment system. Therefore in upgrading situations, emphasis should be placed on the effects of upgrading on the treatment plant as a whole rather than on a particular unit operation.

Various case histories are discussed to illustrate procedures which have been used to upgrade wastewater treatment facilities. Cost estimates for these cases are furnished when such information is available.

13.2 Case History No. 1

Upgrading Using Chemical Addition to Primary Clarifiers and Conversion of Conventional Activated Sludge to Contact Stabilization (1)

Plant 1 before upgrading was a parallel activated sludge and trickling filter plant. The original nominal design capacity of the plant was 3.6 mgd, but the average influent flow had increased to 6.0 mgd. The plant was removing significantly less than 90 percent of the BOD during the summer and as little as 60 percent during the winter.

Upgrading of this plant was an interim measure, since the plant was scheduled to be replaced in 5 years with a regional system. Figure 13-1 is a flow diagram of the upgraded treatment system. Table 13-1 summarizes plant operating conditions prior to upgrading and plant design conditions after upgrading. The conventional activated sludge system originally designed to treat a flow of 1.2 mgd, was upgraded to a contact stabilization system. This was accomplished by increasing the capacity of the mechanical aerators from a total of 40 hp up to 110 hp (60 hp in the contact basin and 50 hp in the stabilization basin) and by modifying the basin's piping. Since experience has shown that contact stabilization works well without primary clarification, the existing primary clarifier was converted to a secondary clarifier to decrease the activated sludge secondary clarifier overflow rate. The contact stabilization process was designed to handle a flow of 3.0 mgd.

The trickling filter portion of the plant was originally designed to operate at 2.4 mgd as a two-stage system. However, during periods of hydraulic overloading, the trickling filters were operated in parallel to increase the hydraulic capacity.

Plant operating data indicated that the secondary portion of the parallel-operated trickling filter system was removing 70-80 percent of the primary effluent BOD. Therefore, it was

FIGURE 13-1
CASE HISTORY NO. 1
FLOW DIAGRAM OF PLANT INCLUDING UPGRADED UNIT PROCESSES

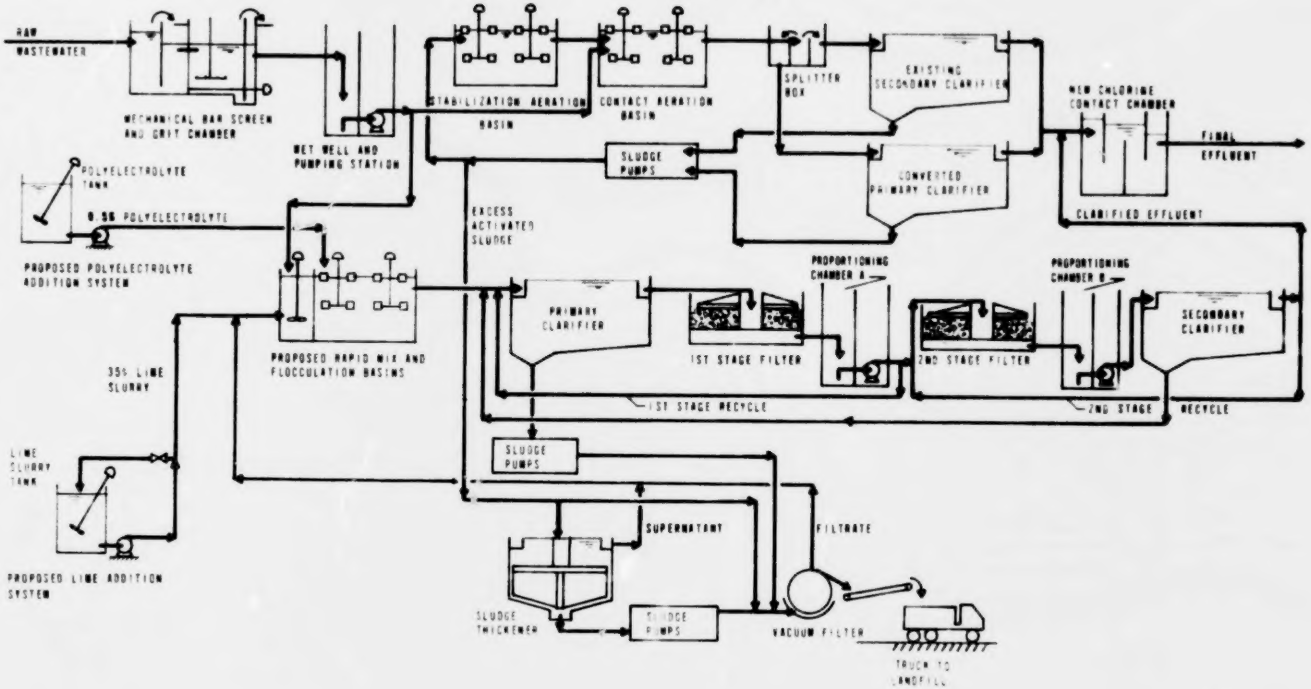

RAW WASTEWATER
MECHANICAL BAR SCREEN AND GRIT CHAMBER
WET WELL AND PUMPING STATION
STABILIZATION AERATION BASIN
CONTACT AERATION BASIN
SPLITTER BOX
EXISTING SECONDARY CLARIFIER
CONVERTED PRIMARY CLARIFIER
NEW CHLORINE CONTACT CHAMBER
FINAL EFFLUENT
SLUDGE PUMPS
CLARIFIED EFFLUENT
POLYELECTROLYTE TANK
0.5% POLYELECTROLYTE
PROPOSED POLYELECTROLYTE ADDITION SYSTEM
EXCESS ACTIVATED SLUDGE
PROPOSED RAPID MIX AND FLOCCULATION BASINS
PRIMARY CLARIFIER
1ST STAGE FILTER
PROPORTIONING CHAMBER A
2ND STAGE FILTER
PROPORTIONING CHAMBER B
SECONDARY CLARIFIER
1ST STAGE RECYCLE
2ND STAGE RECYCLE
35% LIME SLURRY
LIME SLURRY TANK
PROPOSED LIME ADDITION SYSTEM
SLUDGE PUMPS
SUPERNATANT
FILTRATE
SLUDGE THICKENER
SLUDGE PUMPS
VACUUM FILTER
TRUCK TO LANDFILL

Table 13-1

Case History No. 1 - Plant Operating and Design Conditions

Operation	Operating Conditions Before Upgrading	Design Conditions After Upgrading	Remarks
Pre-treatment Facilities (Bar Screen, Aerated Grit Chamber)			
Average Flow, mgd	6.0	6.0	
Activated Sludge Plant			
Primary Clarifier Overflow Rate, gpd/sq.ft.[1]	1,200	--	Primary clarifier converted to secondary clarifier.
Flow to Aeration Basin, mgd	3.0	3.0	
Volumetric Loading in Aeration Tank, lbs. BOD/day/1,000 cu.ft.	–	60[2]	O_2 supply, not organic loading, was limiting prior to upgrading.
Average Sludge Recycle, percent	30	100	
Secondary Clarifier Overflow Rate, gpd/sq.ft.[1]	960	740	Reduction in overflow rate is achieved by converting the primary clarifier to a secondary unit.
BOD Removal in Secondary Treatment, percent	75	90	
Suspended Solids Removal in Secondary Treatment, percent	75	90	
Trickling Filter Plant			
Primary Clarifier Overflow Rate, gpd/sq.ft.[1]	820	820	
BOD Removal in Primary Treatment, percent	20	50 to 60	Chemical coagulation
Flow to Trickling Filter, mgd	3.0	3.0	
Hydraulic Loading, mgd/acre-ft.[3]	--	6.25	For each stage
Organic Loading, lbs. BOD/day/1,000 cu.ft.[3]	--	72	First Stage
		19	Second Stage
Recirculation Ratio	1.0	1.0	For each stage
Secondary Clarifier Overflow Rate, gpd/sq.ft.[4]	1,000	1,000	
BOD Removal in Secondary Treatment, percent	70 to 80	70 to 80	
Sludge Handling Facilities			
Vacuum Filter Operation	16 hrs./day 5 days/week	20 hrs./day 6 days/week	Increased operation to accommodate increased sludge quantities

[1] Clarifier overflow rates based on average flow

[2] Organic loading increases due to increase in flow and elimination of primary treatment.

[3] Does not include recirculation.

[4] Based on average flow plus recycle.

decided to confine the upgrading to improvement of the removal efficiency of the primary clarifier. The selected approach was chemical addition, namely lime and anionic polyelectrolyte. Laboratory testing indicated that as much as 60 percent BOD removal and 75 percent suspended solids removal could be attained in the primary clarifier with the addition of 1.0 mg/l polyelectrolyte and 200 mg/l of lime. With this modification, the trickling filter system could again be operated as a 2-stage system at a flow of 3.0 mgd.

A summary of the measured performance of the existing system and the anticipated performance after upgrading is given below in Table 13-2.

Table 13-2

Summary of Treatment Performance
for Case History No. 1

	Measured Performance Before Upgrading	Anticipated Performance After Upgrading
Design Flow, mgd	3.6	6.0
Average Flow, mgd	6.0	6.0
BOD Removal, percent	80	90
Effluent BOD, mg/l	40	20
SS Removal, percent	85	90
Effluent SS, mg/l	30	20

The cost of upgrading the plant was estimated at $510,000, and the increased annual operating costs were estimated at $74,800. These costs are broken down in Table 13-3.

Table 13-3

Summary of Upgrading Costs
for Case History No. 1
(ENR Index 1500)

ESTIMATED CONSTRUCTION COST

Chemical Addition/Flocculation	$ 45,000
Conversion to Contact Stabilization	340,000
Mechanical Refurbishing	25,000
Electrical Refurbishing	50,000
Construction Contingency	50,000
Total Estimated Construction Cost[1]	$510,000
Unit Capital Cost of Upgrading	$212/1,000 gpd of upgraded flow

ESTIMATED ANNUAL OPERATING COST INCREASE

Operation and Maintenance	$ 5,700
Electrical	9,100
Chemical	52,000
Sludge Disposal	8,000
Total Annual Operating Cost Increase	$ 74,800
Unit Annual Operating Cost Increase	3.4 cents/1,000 gallons of treated flow

[1] Costs do not include engineering, legal, and administrative fees.

13.3 Case History No. 2

Upgrading an Existing High-Rate Trickling Filter Plant Using a Series Activated Sludge Process (2)

Plant 2 before upgrading was a high-rate trickling filter plant whose original design capacity of 12.5 mgd was being approached. The initial treatment requirements were 85 percent BOD and SS removals, and removals of approximately 80 percent were actually being obtained. The treatment requirements were then changed to include 90-95 percent removal of the total oxygen demand (the sum of the ultimate carbonaceous oxygen demand and the nitrogenous oxygen demand). This additional requirement indicated that most of the ammonia nitrogen had to be removed or converted to the nitrate form.

Figure 13-2 is a flow diagram of the upgraded treatment system. Table 13-4 summarizes plant operating conditions prior to upgrading and plant design conditions after upgrading. An in-plant survey indicated that the existing equipment was in good condition and could be incorporated into the upgrading scheme. After an evaluation of alternative upgrading possibilities, the existing plant was upgraded from 12.5 mgd to 22.5 mgd using a series treatment of activated sludge and trickling filtration. The activated sludge system was designed with a detention time of 3 hours, with the activated sludge effluent being subsequently treated in the trickling filter to take advantage of the incipient nitrifying ability of the trickling filter.

Another major consideration in the upgrading was the disposal of waste activated sludge. To reduce the volume of primary and waste activated sludge, a gravity thickener was incorporated into the design. Although the existing two-stage anaerobic digesters had sufficient volume to handle the thickened sludge from the upgraded plant, operational problems were being experienced in the existing plant due to incomplete mixing and non-uniform temperature control in the primary digesters. The original design included mechanical mixing using draft tubes and internal heating coils. To improve digester performance, gas recirculation was used for better mixing, and an external heat exchanger was incorporated to facilitate maintenance of a uniform temperature throughout the primary digesters. When the upgraded plant comes on line, it is expected to perform as indicated in Table 13-5.

FIGURE 13-2

CASE HISTORY NO. 2

FLOW DIAGRAM OF PLANT INCLUDING UPGRADED UNIT PROCESSES

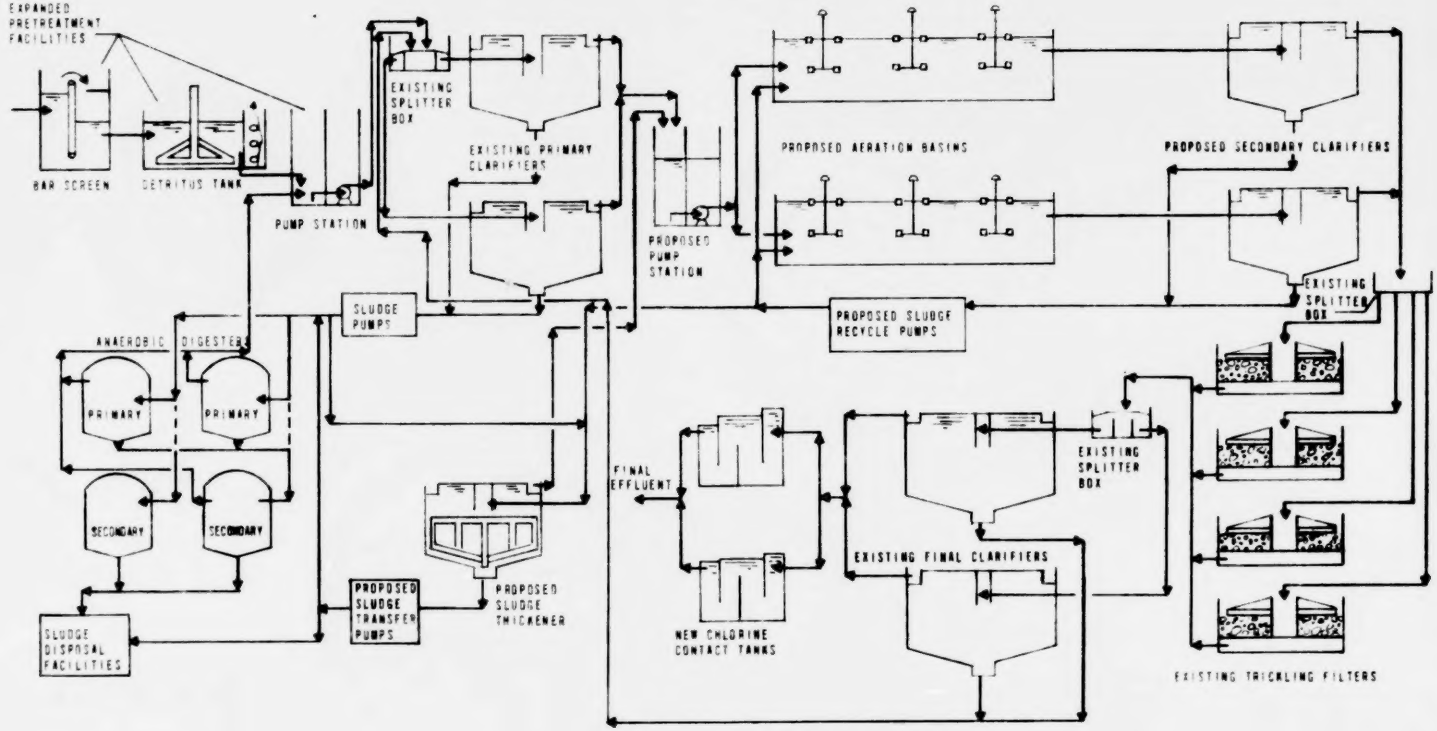

EXPANDED PRETREATMENT FACILITIES
BAR SCREEN
DETRITUS TANK
PUMP STATION
EXISTING SPLITTER BOX
EXISTING PRIMARY CLARIFIERS
SLUDGE PUMPS
ANAEROBIC DIGESTER
PRIMARY
PRIMARY
SECONDARY
SECONDARY
SLUDGE DISPOSAL FACILITIES
PROPOSED SLUDGE TRANSFER PUMPS
PROPOSED SLUDGE THICKENER
PROPOSED PUMP STATION
PROPOSED AERATION BASINS
PROPOSED SECONDARY CLARIFIERS
PROPOSED SLUDGE RECYCLE PUMPS
EXISTING SPLITTER BOX
EXISTING TRICKLING FILTERS
EXISTING SPLITTER BOX
EXISTING FINAL CLARIFIERS
FINAL EFFLUENT
NEW CHLORINE CONTACT TANKS

Table 13-4

Case History No. 2 - Plant Operating and Design Conditions

Operation	Operating Conditions Before Upgrading	Design Conditions After Upgrading
Pre-treatment Facilities (Bar Screen, Grit Chamber, Raw-Sewage Pump Station) Avg. Capacity, mgd	12.5	22.5
Primary Clarifiers		
Overflow Rate, gpd/sq.ft.	766[1]	785[2]
BOD Removal, percent	56	30
SS Removal, percent	74	60
Aeration Basins		
Detention Time, hours	--	3.0
Volumetric Loading, lbs. BOD/day/1,000 cu.ft.	--	30
Organic Loading, lbs. BOD/day/lb. MLVSS	--	0.25
Sludge Recycle, percent of average flow	--	60
Mechanical Aerator Capacity, hp	--	750
Intermediate Clarifiers		
Overflow Rate, gpd/sq.ft.		785
Trickling Filter		
Hydraulic Loading, mgad	30	21.6
Organic Loading, lbs. BOD/day/1,000 cu.ft.	66	21
Recirculation Ratio	1.5[3]	0
Final Clarifiers		
Overflow Rate, gpd/sq.ft.	766[4]	785
Sludge Thickener		
Design Underflow Concentration	--	4.0
Design Solids Loading, lbs. SS/day/sq.ft.	--	10.0
Detention Time, hours	--	10.0
Anaerobic Digesters		
Loading, lbs. VSS/day/cu.ft.	0.05	0.12
Detention Time, days	34	14

[1] Based on average design flow plus 75% secondary sludge recirculation.

[2] Excluding sludge recycle.

[3] The recirculation ratio includes 75% secondary sludge recirculation to the primary clarifier influent and 75% secondary clarifier effluent recirculation to the trickling filter influent.

[4] Based on average design flow plus 75% secondary clarifier effluent recirculation.

Table 13-5

Summary of Treatment Performance
for Case History No. 2

	Before Upgrading	After Upgrading
Average Design Flow, mgd	12.5	22.5
BOD Removal, percent	81	>90
Effluent BOD, mg/l	41	15
SS Removal, percent	81	>90
Effluent SS, mg/l	40	<10
Effluent NH_3-N, mg/l	27	2
Effluent NO_3-N[1], mg/l	<1	26

[1] Approximate

The construction costs for treatment facilities were estimated at $4,130,000, and the increased annual operating costs due to upgrading were estimated at $224,000. These costs are broken down in Table 13-6.

Table 13-6

Summary of Upgrading Costs
for Case History No. 2
(ENR Index 1500)

ESTIMATED CONSTRUCTION COST

Expansion of Pre-treatment Facilities	$ 165,000
Expansion of Primary Treatment	190,000
Addition of Activated Sludge Treatment and Intermediate Clarification	2,030,000
Expansion of Sludge Handling	54,000
Expansion of Control Building, Chlorination Facilities, etc.	260,000
Piping, Electrical, Instrumentation, etc.	890,000
Construction Contingency	541,000
Total Estimated Construction Cost[1]	$4,130,000
Unit Capital Cost of Upgrading	$413/1,000 gpd of upgraded flow

ESTIMATED ANNUAL OPERATING COST INCREASE

Labor	$ 30,000
Power	86,000
Maintenance and Supplies	108,000
Total Annual Operating Cost Increase	$224,000
Unit Annual Operating Cost Increase	2.7 cents/1,000 gallons of treated flow

[1] Costs do not include engineering, legal, and administrative fees.

13.4 Case History No. 3

Use of Roughing Filter to Upgrade an Existing Low-Rate Trickling Filter Plant (3) (4) (5)

Case History No. 3 involves the upgrading of an existing low rate trickling filter plant in Huber Heights, Ohio. The original plant was designed in August, 1956 for a flow of 0.7 mgd, with 85 percent BOD and suspended solids removals. The community developed so rapidly that by 1970, the average flow had increased to 2.3 mgd. The flow diagrams for the original and upgraded plant are shown in Figure 13-3.

Operational and performance data for the overloaded plant for 1962, when the plant was receiving 1.15 mgd, are compared with corresponding data after the plant was upgraded to 2.3 mgd in Table 13-7.

The comminuter and primary clarifiers were replaced with three Hydrasieve units of 1-mgd capacity each. These units are stationary screens capable of removing 20 to 35 percent of the BOD and suspended solids. A Hydrasieve unit is illustrated in Figure 13-4, along with a schematic flow diagram through the unit. These screens generally require no power and little maintenance.

The plastic media roughing filter used in the upgrading has an application rate of approximately 2.5 gpm/sq.ft. Present BOD removal is about 25 to 35 percent through the roughing unit. Because of the increased hydraulic loading, it was necessary to expand the secondary clarification and chlorine contact tank capacities. The abandoned primary clarifiers were converted into sludge thickeners. This step, in addition to conversion of the anaerobic digester to a high-rate unit using gas recirculation for mixing, enabled the sludge handling system to process the increased quantity of sludge produced.

This case history points out the fact that an existing plant may be gradually upgraded to handle a three-fold increase in flow with the use of innovative techniques and newly applied process equipment. The capital costs for upgrading the capacity of the plant were estimated at approximately $300,000 (ENR 1500), or $187 per 1,000 gallons/day of upgraded flow capacity.

FIGURE 13-3

CASE HISTORY NO.3

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS

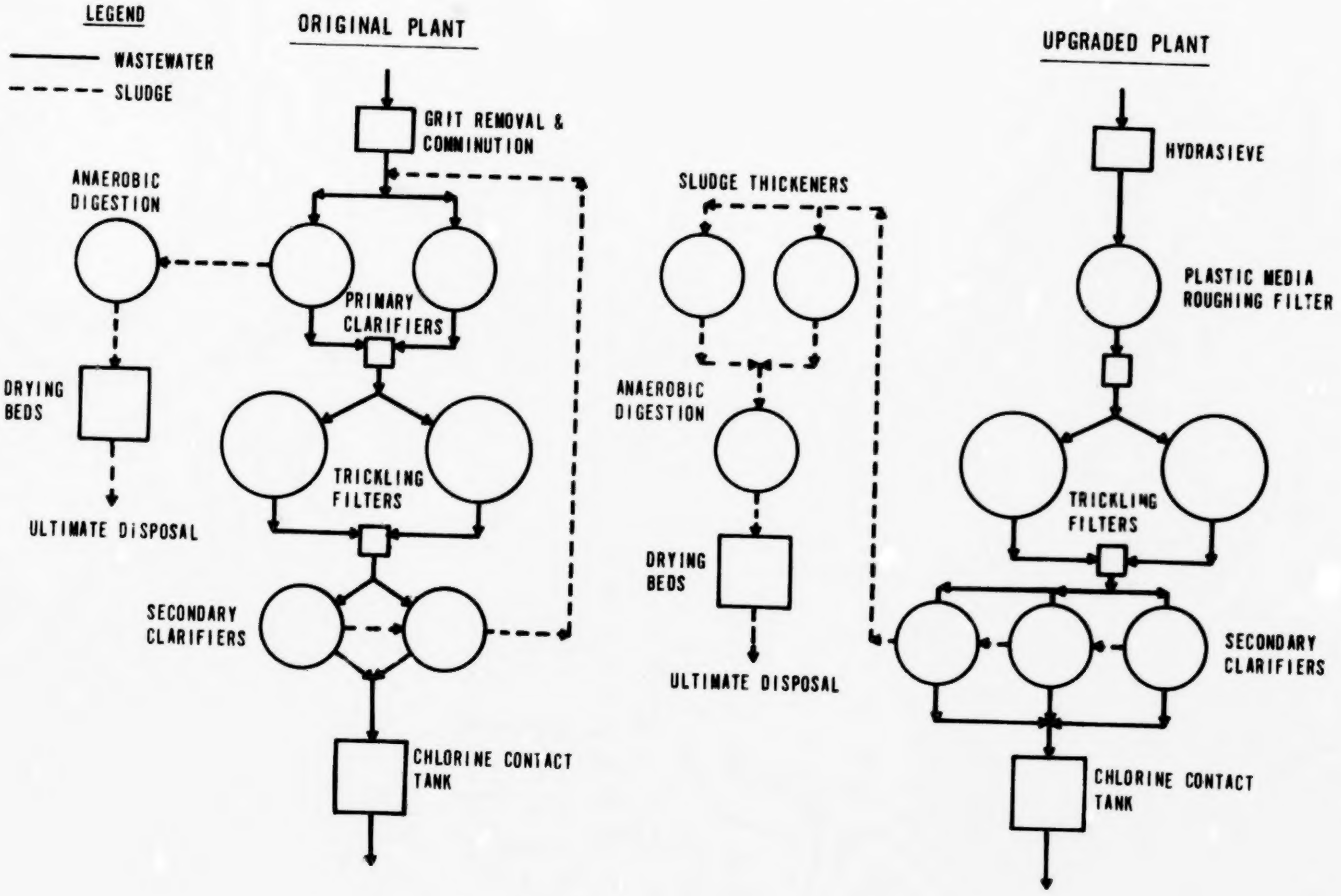

LEGEND
WASTEWATER
SLUDGE
ORIGINAL PLANT
GRIT REMOVAL & COMMINUTION
ANAEROBIC DIGESTION
PRIMARY CLARIFIERS
DRYING BEDS
TRICKLING FILTERS
ULTIMATE DISPOSAL
SECONDARY CLARIFIERS
CHLORINE CONTACT TANK
SLUDGE THICKENERS
ANAEROBIC DIGESTION
DRYING BEDS
ULTIMATE DISPOSAL
UPGRADED PLANT
HYDRASIEVE
PLASTIC MEDIA ROUGHING FILTER
TRICKLING FILTERS
SECONDARY CLARIFIERS
CHLORINE CONTACT TANK

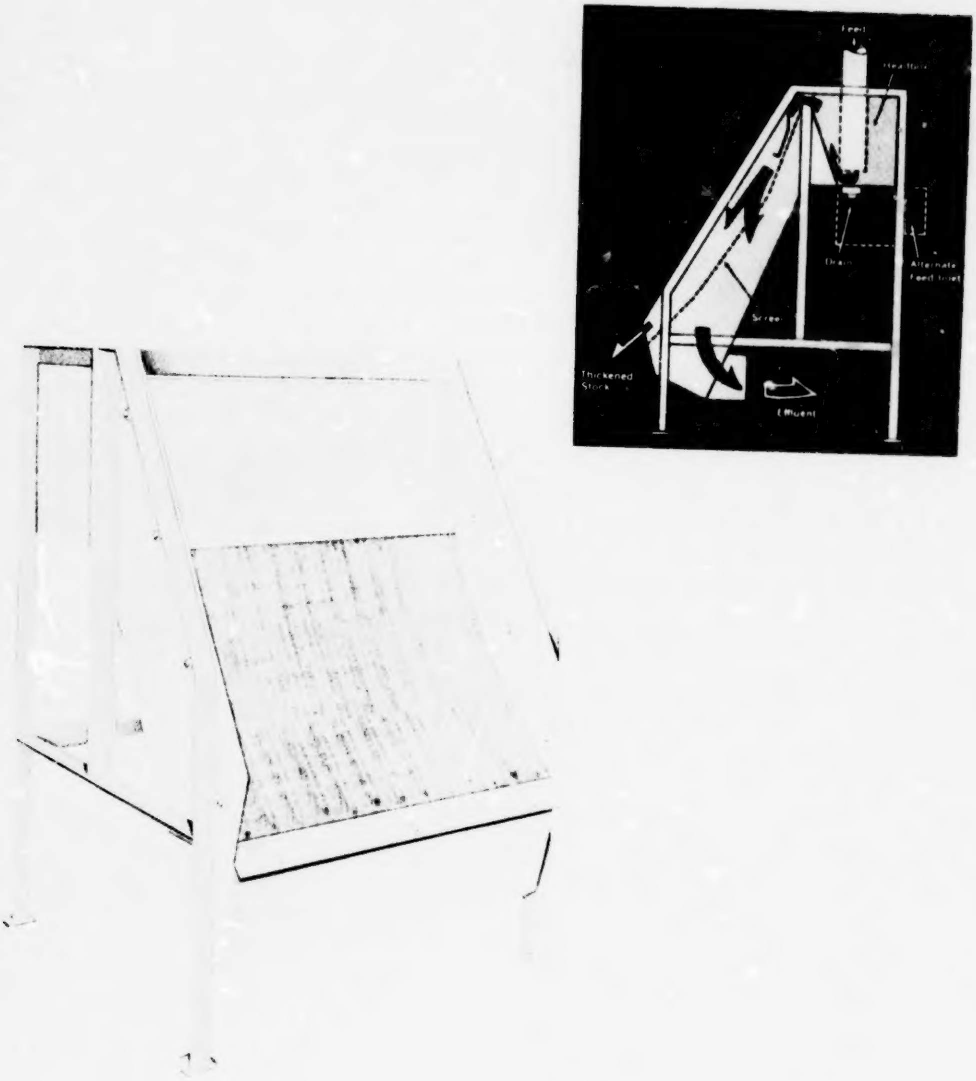
Table 13-7

Case History No. 3 - Plant Operating and Performance Data

Operation	Before Upgrading (1962)	After Upgrading
Average Flow Rate, mgd	1.15	2.3
Primary Clarifier		
Overflow Rate[1], gpd/sq.ft.	1,170	–[5]
Hydrasieve Slot Size, inches	–	0.06
BOD Removal, percent	35[2]	25[3]
SS Removal, percent	–[6]	22[3]
Plastic Media Roughing Filter		
Hydraulic Loading[4], gpm/sq.ft.	–	2.5
Organic Loading[4], lbs. BOD/day/1,000 cu.ft.	–	520
Recirculation Ratio	–	≈2.0
BOD Removal, percent	–	30
Trickling Filter (Stone Media)		
Hydraulic Loading, mgad	6.0	12.0
Organic Loading, lbs. BOD/day/1,000 cu.ft.	56.2	87.0
Secondary Clarifiers		
Overflow Rate, gpd/sq.ft.	1,170	750
Overall Plant Performance		
BOD Removal, percent	83	85
Effluent BOD, mg/l	41	37
SS Removal, percent	–[6]	84
Effluent SS, mg/l	–[6]	40

[1] Based on average flow rate.
[2] Based on primary clarifier performance.
[3] Based on hydrasieve performance only.
[4] Including recirculation.
[5] Primary clarifiers converted to gravity thickeners.
[6] Operating data not available.

FIGURE 13-4
HYDRASIEVE SCREENING UNIT*


Feed
Headbox
Drain
Alternate Feed Inlet
Screen
Thickened Stock
Effluent

---

*COURTESY OF THE BAUER BROS. CO. – SPRINGFIELD, OHIO

13.5 Case History No. 4

Upgrading an Existing High-Rate Trickling Filter by Conversion to a Super-Rate Filter System (6)

The North Treatment Plant at Sedalia, Missouri, is a high-rate trickling filter plant, designed for 1.25 mgd, and was removing 85 percent of the BOD in 1963. However, the Water Pollution Board set a final effluent BOD of 20 mg/l, which Sedalia could not meet with the existing facilities. The 1963 plant flow diagram is illustrated in Figure 13-5.

The plant was upgraded to treat an average design flow of 2.5 mgd. The existing stone-media filter was renovated to operate with plastic media. In addition, a second plastic-media filter was constructed. The two plastic-media filters are operated in parallel with a total recirculation ratio of 1.55. One additional primary clarifier and one additional secondary clarifier were installed.

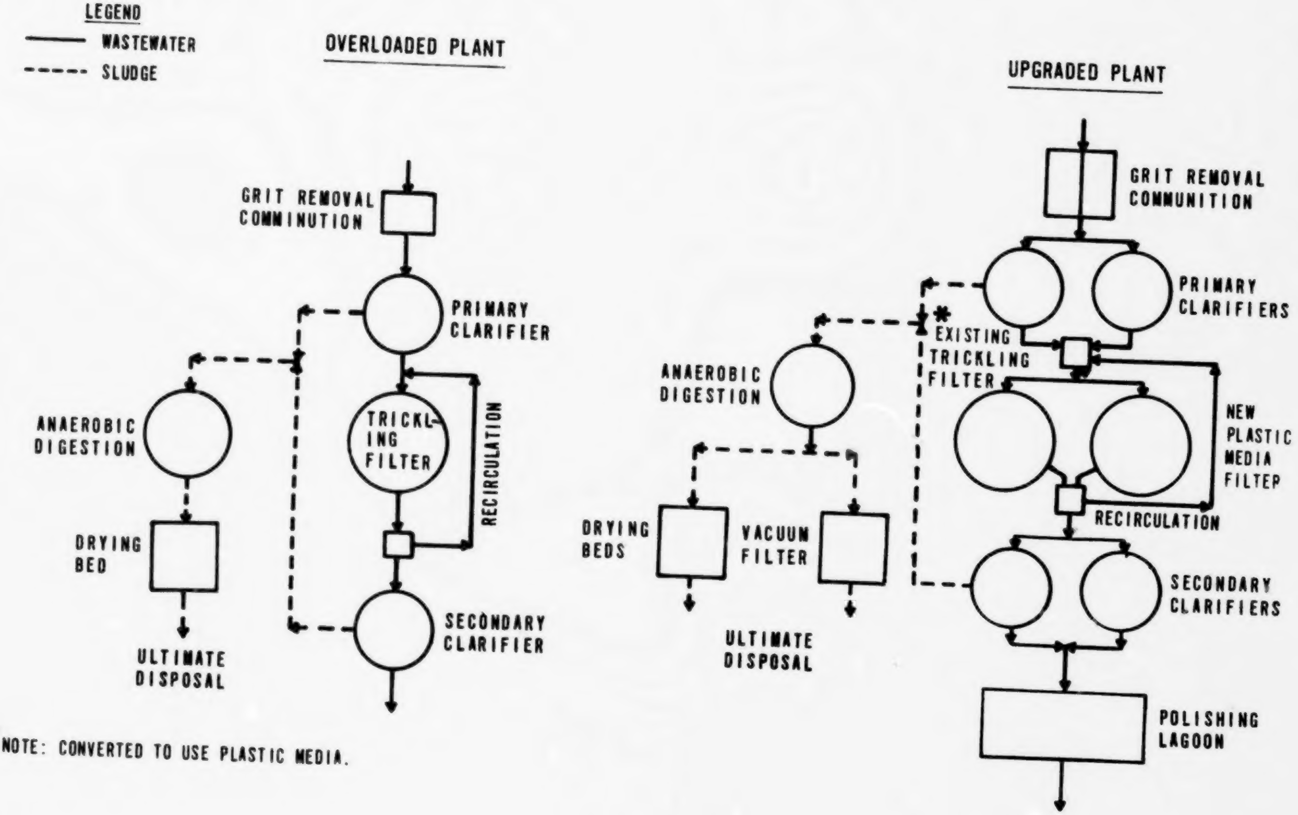
To remove additional BOD and suspended solids, a shallow aerobic polishing lagoon was constructed after the secondary clarifiers, and a vacuum filter was added to reduce the volume of digested sludge. A flow diagram of the upgraded plant is also shown in Figure 13-5. Table 13-8 contains a summary of operating data for the 1963 overloaded period; in addition, the upgraded design criteria are listed, along with actual operational data for the post-upgrading period. It should be noted that the effluent BOD was improved from 115 mg/l to 11 mg/l after upgrading, which is below the 20 mg/l requirement. It should also be pointed out that Missouri has no suspended solids removal requirements for plants with a flow of less than 10 mgd.

The capital costs of upgrading the plant were estimated at $2,600,000 (ENR Index 1,500), or $2,080 per 1,000 gpd of incremental upgraded capacity.

FIGURE 13-5

CASE HISTORY NO. 4

COMPARISON OF ORIGINAL AND UPGRADED FLOW DIAGRAMS


LEGEND
WASTEWATER
SLUDGE
OVERLOADED PLANT
GRIT REMOVAL COMMINUTION
PRIMARY CLARIFIER
TRICKLING FILTER
RECIRCULATION
SECONDARY CLARIFIER
ANAEROBIC DIGESTION
DRYING BED
ULTIMATE DISPOSAL
UPGRADED PLANT
GRIT REMOVAL COMMUNITION
PRIMARY CLARIFIERS
* EXISTING TRICKLING FILTER
NEW PLASTIC MEDIA FILTEP
RECIRCULATION
SECONDARY CLARIFIERS
POLISHING LAGOON
ANAEROBIC DIGESTION
DRYING BEDS
VACUUM FILTER
ULTIMATE DISPOSAL

*NOTE: CONVERTED TO USE PLASTIC MEDIA.

Table 13-8

Case History No. 4 - Plant Operating and Design Data

Operation	1963 Operating Data	Upgraded Design	1969 Operating Data
Average Daily Flow, gpm	1.25	2.5	1.80
Raw Wastewater BOD, mg/l	768	576	450
Primary Clarifiers			
Overflow Rate[1], gpd/sq.ft.	1,000	1,000	720
BOD Removal, percent	40	40	60
Trickling Filters			
Hydraulic Loading[2], mgad	24.5	31.8	25
Organic Loading[2], lbs. BOD/day/ 1,000 cu.ft.	226	72.6	20
Recirculation Ratio	1.0	1.55	1.55
Final Clarifiers			
Overflow Rate[1], gpd/sq.ft.	755	755	545
Secondary BOD Removal, percent	75	93.2	86.8
Polishing Lagoon (Shallow Aerobic)			
Maximum BOD Loading, lbs. BOD/ acre/day	–	68	30
BOD Removal, percent	–	12	54
Vacuum Filtration Rate, lbs./sq.ft./hour	–	5.0	–[3]
Overall Plant Performance			
BOD Removal, percent	85	96.5	97.7
Effluent BOD, mg/l	115	20	11

[1] Based on average daily flow.
[2] Including Recirculation.
[3] Lack of sufficient operating data.

13.6 Case History No. 5

Upgrading a Contact Stabilization Package Plant to a "Modified" Completely-Mixed Flow Pattern (7)

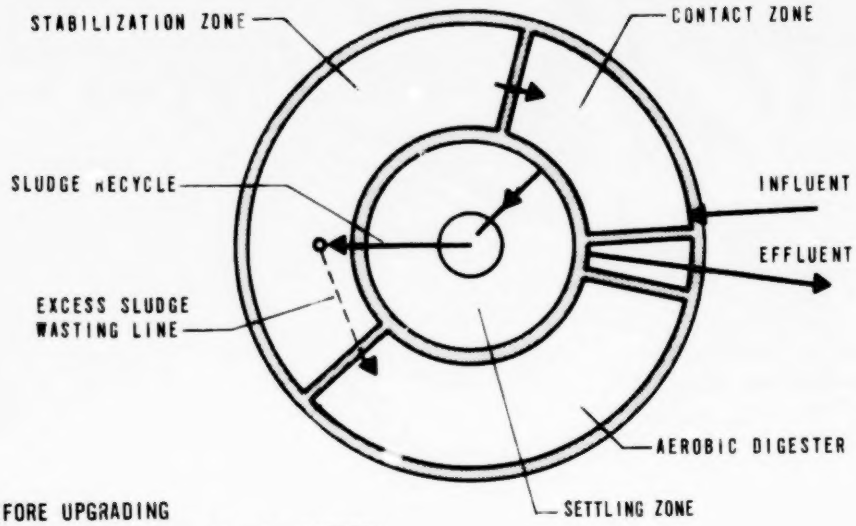
Coralville, Iowa, had a contact stabilization package plant which was providing detention times of 2.6 and 6.5 hours, respectively, in the contact and stabilization zones, based on a forward flow of 867,000 gallons per day, not including sludge recycle. A typical plan view of the contact stabilization package plant is shown in Figure 13-6.

As previously discussed in Chapter 5, a contact-zone detention time of this magnitude may result in improper plant operation because the sludge becomes partially stabilized in this zone and acquires poorer settling characteristics. After investigation, this was found to be the case in Coralville. Operational data from the plant before upgrading are summarized in Table 13-9. The effluent BOD and suspended solids were 26 and 24 mg/l, respectively.

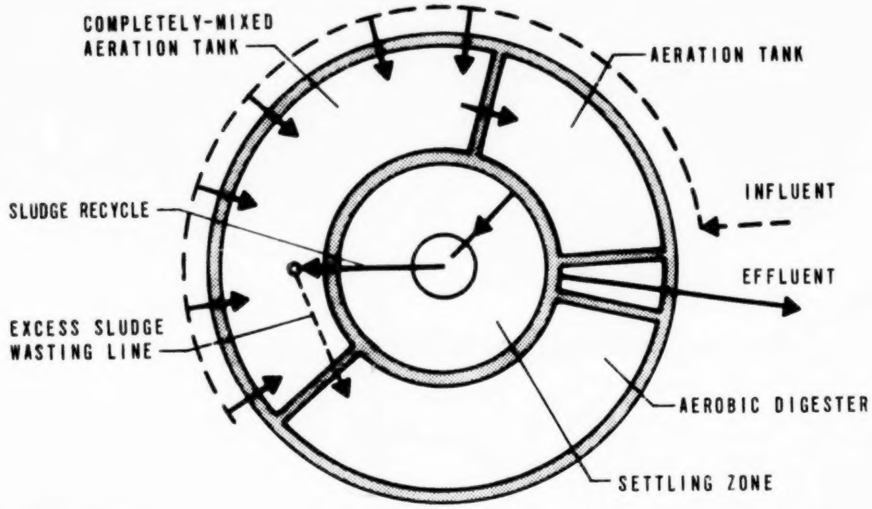
To improve the plant's performance, it was decided to modify the flow pattern as indicated in Figure 13-6. The influent piping was modified so that the raw wastewater was evenly distributed into what originally was the stabilization zone. No wastewater was introduced into the former contact zone. Mixed liquor in the upgraded system proceeded from the former stabilization zone through the former contact zone to the secondary clarifier. The return sludge was introduced into the former stabilization zone at one point only. Therefore, the upgrading resulted in a "modified" completely-mixed flow pattern, with an overall detention time of 9.1 hours for an average flow of 867,000 gpd.

Performance data for the upgraded plant are included in Table 13-9. The effluent BOD and suspended solids concentrations were lowered to 13 and 6 mg/l, respectively, by the upgrading procedure. The costs associated with this modification are primarily due to piping changes. No cost breakdown was available for the modification.

FIGURE 13-6
CASE HISTORY NO. 5
UPGRADING A CONTACT STABILIZATION PACKAGE PLANT TO A COMPLETELY-MIXED FLOW PATTERN (5)


STABILIZATION ZONE
CONTACT ZONE
SLUDGE RECYCLE
INFLUENT
EFFLUENT
EXCESS SLUDGE WASTING LINE
AEROBIC DIGESTER
SETTLING ZONE

BEFORE UPGRADING
CONTACT STABILIZATION FLOW PATTERN


COMPLETELY-MIXED AERATION TANK
AERATION TANK
INFLUENT
SLUDGE RECYCLE
EFFLUENT
EXCESS SLUDGE WASTING LINE
AEROBIC DIGESTER
SETTLING ZONE

AFTER UPGRADING
MODIFIED COMPLETELY-MIXED FLOW PATTERN

Table 13-9

Case History No. 5 - Plant Operating Data

Operation	Contact Stabilization[1]	Modified Completely-Mixed[2]
Average Flow, mgd	0.867	0.867
Influent BOD, mg/l	135	135
Influent SS, mg/l	150	150
Aeration Tank		
Sludge Recycle, percent	60	60
Contact Unit Volumetric Load, lbs. BOD/day/1,000 cu.ft.	78	22[4]
Contact Unit Organic Load, lbs. BOD/day/lb. MLVSS	0.4	–
Contact Unit MLSS, mg/l	3,500	–
Secondary Clarifier		
Overflow Rate[3], gpd/sq.ft.	750	750
Overall Plant Performance		
BOD Removal, percent	81	90
Effluent BOD, mg/l	26	13
SS Removal, percent	84	96
Effluent SS, mg/l	24	6

[1] Before Upgrading
[2] After Upgrading
[3] Based on average flow
[4] Based on total aeration volume.

13.7 Case History No. 6

Upgrading Using Polyelectrolyte Addition Before the Primary Clarifier (8)

The Easterly Wastewater Treatment Works in Cleveland, Ohio, is a conventional activated sludge plant whose dry-weather design flow is 123 mgd. In 1968, the plant was hydraulically overloaded. Operational data from the period are presented in Table 13-10, identified as the 1968 Control Period. It is interesting to note that the overflow rate of the primary clarifiers was 2,030 gpd/sq.ft., which is substantially above the accepted values. As a result of the hydraulic overloading, the overall BOD and suspended solids removals were only 79 and 85 percent, respectively.

To improve the overall plant performance, a polyelectrolyte addition program was initiated. An anionic polyelectrolyte, Purifloc A-23, was added at an average dosage of 0.21 mg/l. Since proper polyelectrolyte dispersal and uniform mixing into the entire waste flow constitute an extremely important aspect of the flocculation process, it was decided to add the polymer at the plant's two Venturi meters. These meters are located immediately after the grit chamber and in front of the pre-aeration basin. Dye studies indicated that there was a 7.5-minute travel time between the Venturi meters and the primary clarifiers. Six of the 7.5 minutes would be spent in the pre-aeration basin. The gentle agitation in the pre-aeration basin provided adequate flocculation of wastewater solids.

A summary of the effectiveness of the polymer addition on plant performance as compared to performance during the previous control period is also presented in Table 13-10. The improvement in primary suspended solids removal increased the volume of primary sludge from 5.0 to 6.8 million gallons per month. Overall plant performance was improved, as noted by the reduced effluent BOD and suspended solids concentrations. In addition to the increased treatment efficiency, the polyelectrolyte addition was responsible for the following benefits to the subsequent activated sludge process:

1. A 20 percent volume decrease in the amount of waste activated sludge produced.
2. A 22 percent reduction in air supply requirements, resulting in a power cost savings of over $3,300 per month.
3. An increase in aeration tank D.O. concentration from an average of 3.2 mg/l to 3.8 mg/l.

An economic comparison was made between polyelectrolyte addition and providing additional tankage to equal the performance of the flocculation system The amortized cost for the additional tankage was about $314,000 per year, while the chemical cost was $150,000 per year, thereby indicating a considerable yearly savings in favor of the polyelectrolyte alternative.

Table 13-10

Case History No. 6 - Plant Operating Data

Operation	1968 Control Period	Polymer Demonstration Period
Influent BOD, mg/l	104	67
Influent SS, mg/l	169	157
Primary Clarifier		
Overflow Rate, gpd/sq.ft.	2,030	2,170
BOD Removal, percent	31	46
SS Removal, percent	31	51
Sludge Solids Concentration, percent	4.1	4.3
Sludge Volume Pumped, million gallons/month	5.0	6.8
Aeration Tank		
MLSS, mg/l	1,670	1,602
Organic Loading, lbs. BOD/day/lb. MLSS	0.48	0.29
Dissolved Oxygen, mg/l	3.2	3.8
Waste Activated Sludge Concentration, percent	2.4	2.0
Waste Activated Sludge Pumped, million gallons per month	12.3	9.8
Overall Plant Performance		
BOD Removal, percent	79.1	83.4
Effluent BOD, mg/l	21.8	11.1
SS Removal, percent	85.3	89.2
Effluent SS, mg/l	24.8	17.0

13.8 References

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CHAPTER 14

OPERATION AND MAINTENANCE REQUIREMENTS FOR UPGRADED PLANTS

14.1 General

The discussions in the preceding chapters of this manual demonstrate the fact that upgrading existing wastewater treatment facilities involves the consideration and possible inclusion of a multiplicity of processes and process equipment. Operation and maintenance requirements for a plant incorporating such a diversity of processes are frequently complex. Since this manual is directed principally to the process design of upgraded facilities, it is not appropriate to the scope or purpose of this document to present a detailed operation and maintenance manual. Rather, some guidelines and pertinent discussion will be presented to assist in the development of an operation and maintenance program for an upgraded plant.

14.2 Responsibility of the Design Engineer

One of the first and more obvious considerations to the engineer faced with upgrading existing wastewater treatment facilities is the need to consider current plant operation and maintenance, either as causes of problems requiring upgrading or as means of accomplishing the required improvement in plant performance. As has been discussed in the preceding chapters dealing with individual unit processes and operations, an important first step in upgrading plant performance is to review operational and maintenance practices, and the adequacy and competency of plant operating staff and supervision. It is frequently possible to effect a significant improvement in plant performance, without the need for capital expenditures, by modification of plant practices, by addition of operating and/or maintenance staff, by improved or supplemented supervision, or by training of existing staff.

In September, 1970, the Federal Water Quality Administration (now EPA-OWP) issued "Federal Guidelines - Design, Operation and Maintenance of Wastewater Treatment Facilities" (1). The latter section of this document includes "Guidelines for Operation and Maintenance". Projects for which Federal grant assistance is requested are expected to comply with these guidelines as well as with technical bulletins which the Environmental Protection Agency will be issuing from time to time. The aforementioned guidelines provide information on the basic minimum requirements for federally assisted projects on such matters as personnel, records, reports, laboratory control, and process control, and present a suggested guide for an operation and maintenance manual.

The design engineer has the responsibility to be fully knowledgeable of the operational factors of the plant for which he is responsible and to relate that knowledge to the alternative means of upgrading the plant's performance. For instance, in both the selection and evaluation of alternative upgrading measures, he must consider total annual costs, including operation and maintenance, as well as capital construction costs. Some guidelines to assist in such comparisons have been included in the previous discussions on unit processes.

The design engineer should make it a point to obtain all pertinent plant operating records and to solicit the observations, comments, and suggestions of the operating staff so that optimum use may be made of the experience and knowledge of the particular plant and system. He should also prepare or assist in the preparation of operation and maintenance manuals for the upgraded plant, and in the on-site training of operating, maintenance and supervisory personnel to ensure that the function, capabilities, and limitations of the upgraded facilities are adequately communicated to those responsible for their operation. He should be retained to remain involved during at least the initial operational period to utilize most effectively his expertise and knowledge of the process.

The design engineer also has the responsibility to assess the adequacy of available information on existing plant performance, to define the additional data required, and to develop and direct or execute a program to obtain such information. A program to obtain adequate design information can range anywhere from simple qualitative observations through quantitative samplings and analysis programs to laboratory or pilot-scale treatability studies. This charge presents a very real challenge to his professional judgement, requiring the balancing of the constraints of economics, timing, and staff availability against the need for a fully reliable and effective treatment facility. The information presented in this manual provides direction and guidance to assist the design engineer in the exercise of this professional judgement.

14.3 Instrumentation and Automatic Operation

Historically, wastewater treatment plants have not incorporated sophisticated instrumentation or automated operation. However, with the advent of improved instrumentation and the need in many cases to consistently maintain plant effluent quality, the use of such systems may be indicated.

14.3.1 Instrumentation

In general, instrumentation leads to improved performance, increased efficiency, and reduced operating costs. It also offers permanent records for system evaluation, and generates data for required State regulatory reports. There are three basic types of parameters which may be monitored in wastewater treatment: mechanical, physical, and chemical. Table 14-1 contains a summary of instrumentation available for use in wastewater treatment plants (2). Some common on-line instruments which have application for various unit processes in a conventional wastewater treatment processes follow.

Table 14-1

Wastewater Treatment Instrumentation

Primary Measuring Method	Metering Function	Some Applications
Pressure	Pressure Level	Pump discharge pressure Force main pressure Wet-well level Chemical storage level High-water alarm Gas storage pressure
Mechanical position or motion	Level Position	Digester cover level Valve position indication High-water alarm
Electric and electronic	Temperature Flow Level	Incinerator temperature Wet-well overflow Diversion overflow Digester temperature Magnetic flow metering
Photoelectric	Turbidity Total carbon analyzer	Final effluent turbidity BOD
Differential pressure	Flow Ratio control	Wastewater flow Air flow Cl_2 flow Treatment chemical flow Flow proportioning
Radiation absorption	Specific gravity Density Moisture content Level	Sludge density control Wet-well level Sludge-well level Filter cake moisture content
Electrochemical reaction	D.O. pH ORP Cl_2 concentration Conductivity	Aeration tank D.O. Digester pH Final effluent pH Cl_2 residual Effluent D.O.

Source: Salvatorelli, JWPCF (2)

14.3.1.1 Bar Screen Control

Instrumentation for bar screen control should include an inlet level indicator which actuates the cleaning mechanism when the water reaches a predetermined level in the influent chamber. In addition, a high water surface level alarm and a high torque alarm in the bar screen cleaning mechanism should be considered.

14.3.1.2 Grit Chamber Control

Grit chamber controls should include a high torque alarm on motor controls and perhaps a timer on grit removal mechanisms. Both the bar screen and grit chamber high torque alarms may be connected to a high torque motor shut down.

14.3.1.3 Raw Wastewater Flow Measurements

Some means of measuring wastewater flow include flumes, venturi meters, orifice plates, and magnetic flow meters. Flow measurement is required to provide a record of influent flow to evaluate plant and unit process efficiencies and to dispense chemicals.

14.3.1.4 Primary Clarification Control

One method of measuring primary sludge is with a magnetic flow meter. Self heating meters are available and quite useful where grease is a problem. Sludge depth can be indicated using either infra-red adsorption, photo-electric opacity, ultra-sonic amplitude detection, or radiation amplitude detection. Sludge pumping can then be initiated automatically and controlled by a predetermined timing operation or by using a sludge density meter in the clarifier underflow line.

14.3.1.5 Secondary Clarification Control

The motor controls of the secondary clarifier, like the primary clarifier, should be connected to a high torque alarm and motor shut down switch. Control of secondary sludge drawoff must be considered in conjunction with subsequent sludge dewatering equipment. The control of secondary sludge is much more sensitive than control of primary sludge.

Sludge blanket depth can be determined by the same means as used to determine primary sludge depth. Present measuring and sensing equipment is not as sensitive as required for most secondary sludges. A measurement of sludge blanket depth with a predetermined timed sludge drawoff may prove to be the optimum instrumentation. Control of recycled sludge is probably best accomplished by utilizing a magnetic flow meter as the flow element. A conventional flow element can be used, but would be recommended only if supplemented with a continuous purge system, inspection access, and a good maintenance program. The control loop should include the flow element and controller with proportional and reset controls. Waste sludge should likewise be metered and recorded.

14.3.1.6 Dissolved Oxygen Control (Activated Sludge)

Reasonably reliable and accurate dissolved oxygen monitoring equipment has been developed and field demonstrated in recent years. Automatic control of dissolved oxygen (D.O.) in activated sludge plants is generally more useful and justifiable on the basis of economics of operation rather than on reliability or quality of performance. That is, automatic D.O. control can be used to more effectively match power input to oxygen demand than to upgrade biological removal kinetics. Placement of the D.O. probe to obtain a true average reading is a difficult problem Assistance from the probe manufacturer should be requested if necessary.

14.3.1.7 Trickling Filtration Systems

Trickling filtration generally requires less sophisticated instrumentation than does activated sludge. The instrumentation for trickling filters generally includes control of effluent and sludge recycle depending on the individual flow diagram. Control versatility must be provided if a two-stage plant is to be operated in series and parallel.

14.3.1.8 Chlorination Systems

The chlorination system is generally controlled with a final residual analyzer. The system should include recording as well as control equipment.

14.3.1.9 Sludge Handling Systems

Instrumentation packages for sludge handling systems, including air flotation, anaerobic digesters, vacuum filters, and centrifuges, are often supplied by the individual manufacturer. Coordination is required to couple control to input and output systems, and to relay required information to a central control panel, if used. Detailed discussion of the specific instrumentation involved is omitted here because of the many approaches which may be applicable.

14.3.1.10 Level Control

Level control may be critical in the operation of various unit processes. High and low level alarm may be coupled with pump control as required. Pneumatic bubbler sensing systems offer the simplest approach to level control. A water system that creates an air pressure suitable for a level sensor is now available and is quite useful in systems where there is no instrument air available.

14.3.1.11 Foam Detection

Foam levels can be detected by use of photoelectric cells which use a modulated output. The output from the foam detection device can then actuate foam control equipment.

14.3.2 Automated Instrumentation

Presently there are automated wet chemical procedures for COD, ammonia, phosphates, and other specific materials. In addition, combustion procedures for total carbon (TC), total organic carbon (TOC), and total oxygen demand (TOD) have been developed and applied successfully. These automated analytical systems provide levels of sensitivity which are not normally obtained in manual operation of the same procedures, and in some cases can be used for process control.

14.3.3 Centralized Control Panel

The use of a centralized control panel should be considered in any fully instrumented system. Instant data availability at a central control point permits immediate reaction to alarm conditions and, with proper manual standby controls, gives the experienced operator complete control over most situations.

14.3.4 Summary

The decision to utilize sophisticated instrumentation and/or automatic controls depends, among other factors, on plant size and the importance of consistently maintaining effluent quality.

There is a scarcity of information regarding the current investments made in controls for wastewater treatment plants. Salvatorelli (2) reported that approximately 4.5 percent of the total capital cost of a recently constructed municipal wastewater facility was invested to completely monitor and control the facility. A somewhat analogous comparison is the 12 percent invested by chemical and petroleum industries in 1968 for manufacturing process controls (3). Andrews (3) reported that, based strictly on an economic analysis, a 9 percent investment in instrumentation and controls for a 100 mgd activated sludge plant could be justified if the plant performance could be improved from 87 to 92 percent BOD removal. Smaller plants (less than 5 mgd) may not be able to justify expenditures for such instrumentation and controls even if plant performance could be improved as above. The possibility should be analyzed carefully and weighed against other alternatives, e.g., better use of the money through additional treatment capability.

14.4 Operation and Maintenance Requirements

Operation and maintenance requirements (and their associated costs) are generally defined by three factors: chemicals, required power, and labor.

14.4.1 Chemicals

Chemical quantities and their associated annual costs can be readily estimated from required dosage rates (average for annual costs, maximum for storage and feed rate requirements), wastewater flow rates (average and maximum considerations as with dosage), and unit costs appropriate to the particular chemicals used. The designer should bear in mind that the form in which the chemical is received, stored, or fed can have a significant impact on the amount of operating labor required in its handling and feeding, as well as on the initial construction costs of the handling and feeding facilities. This is an individual consideration requiring evaluation of relative costs as delivered for the various forms, the availability of existing storage, handling, and feeding facilities, and staffing policies.

14.4.2 Power

Incremental power requirements and costs are likewise relatively simple to estimate, being the function of additional connected horsepower, its operating frequency and duration, and unit power costs. Local utility policy with respect to demand rates, of course, must be included in the economic comparison. In addition, recent recommendations by EPA state that the treatment facility should be capable of satisfactory operation during emergencies and power failures (1). To achieve this degree of reliability, duplicate sources of power for essential plant elements must be provided.

14.4.3 Labor

EPA will be issuing recommended wastewater treatment plant staffing guides in the near future. These guides will offer the engineer a means of more accurately estimating manpower requirements by establishing a step-by-step methodology for itemizing those tasks which must be performed at treatment plants, and for applying related manpower needs to these tasks. One step further is the need for employing personnel who are qualified to adequately and efficiently operate the wastewater treatment plant. Most states currently have mandatory operator certification programs, while the majority of the others have voluntary operator certification programs.

14.4.3.1 Operating Manpower Requirements

The first step in the development of estimates of additional operating manpower requirements for upgraded facilities is a thorough analysis of existing operations to establish the adequacy of present staff and operational procedures. This must be done whether the upgrading procedure selected involves the modification of operating procedures or requires the addition or modification of unit processes. The analysis should account for specific local conditions, problems, and objectives.

Following this analysis of existing wastewater treatment facilities, the engineer must make a judgement as to the efficiency of operations and the qualifications of the staff in relation to performance of upgraded facilities. Factors effecting such a judgement include:

1. The physical arrangement and geographic dispersion of the system.

2. The relative complexity of the present versus the upgraded facilities.

3. The level of training and experience of the present staff in relation to the level required by upgraded processes.

4. The relative degree of laboratory and analytical control required for the upgraded versus the existing facilities.

The operational analysis and estimate of manpower requirements should be done on an individual basis for each plant being considered for upgrading. Manpower requirements should be estimated and recommended as dictated by the specific local conditions.

14.4.3.2 Maintenance Manpower Requirements

Similar to the development of estimates for additional operating manpower requirements for upgrading facilities, the first step in the estimation of additional maintenance manpower requirements is a thorough analysis of existing maintenance staff and procedures. The development and implementation of an appropriate preventive maintenance program for wastewater treatment facilities alone can represent a significant upgrading procedure to the extent that it provides for reliable, consistent plant operation and performance. Again, the design engineer responsible for facility upgrading must exercise his judgement as to the sufficiency of maintenance programs and the adequacy of the qualifications and performance of the maintenance staff. This information must in turn be related to the requirements of the upgraded processes and facilities. The factors influencing the engineer's judgement on these aspects of facility performance are similar to those discussed previously on the topic of operation requirements.

In the estimation of total maintenance costs for upgraded facilities, the engineer must include maintenance materials and supplies in addition to maintenance labor. Again, he is usually dealing with incremental costs and must assess the adequacy of present lubrication practices, spare parts stocking and replacement practices, and similar maintenance procedures involving expendable materials and supplies.

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27 *Abstract*

The main purposes of this manual are to examine situations that necessitate upgrading of existing municipal wastewater treatment plants and to discuss and evaluate the corrective actions that are required to upgrade these existing plants. Upgrading to overcome organic and hydraulic overloadings and/or to meet more stringent treatment requirements is considered. The information presented in this manual is specifically adapted to plants having capacities of less than 5 mgd. This particular capacity was selected because most of the existing municipal wastewater treatment plants in the United States have capacities of less than 5 mgd. The manual emphasizes that operational improvement and modifications to existing unit operations be considered as the logical initial approach to upgrading existing treatment plants, before major expansion of existing facilities is implemented. Because of the numerous alternatives available for upgrading an existing treatment plant, it is necessary to understand thoroughly the fundamentals of the various unit operations commonly used in municipal wastewater treatment plants. Therefore, this manual examines in depth the capabilities, limitations, and interrelationships of the various unit processes. The manual also examines hypothetical situations requiring upgrading of unit operations and describes "order of magnitude" costs associated with the upgrading of various unit operations. One chapter of the manual presents case histories of upgrading of existing wastewater treatment plants to illustrate the approaches actually used in these circumstances. The operation and maintenance requirements of the upgraded treatment plants are also briefly examined in the manual.

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